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Pilot Scale Investigation

OF A

VENTURI-TYPE CONTACTOR FOR REMOVAL OF SO_2 BY THE LIMESTONE WET-SCRUBBING PROCESS

PREPARED BY

Cottrell Environmental Systems, Inc.

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Final Report

PILOT SCALE INVESTIGATION OF A VENTURI-TYPE CONTACTOR FOR REMOVAL OF SO_2 BY THE LIMESTONE WET-SCRUBBING PROCESS

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For The

AIR POLLUTION CONTROL OFFICE
OF THE ENVIRONMENTAL PROTECTION AGENCY
Division of Control Systems

Under Contract EHS-D-71-24

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SUMMARY

Control of sulfur dioxide emission from a coal-fired power generating boiler using a cocurrent venturi-type scrubber in series with a wetted film packed tower has been studied in a one-thousand cfm pilot system.

Sulfur dioxide absorption characteristics were studied in detail with three types of alkali materials, calcium oxide, sodium carbonate and calcium carbonate. Sulfated lime/fly ash and dolomitic lime were also tested and their absorption properties were compared to the calcium oxide results.

The primary objectives of this work were the development of design data for predicting sulfur dioxide absorption in 1) a venturi scrubber with limestone-injection wet scrubbing and 2) a combination of a venturi scrubber and packed tower with direct lime/limestone wet scrubbing. A simplified method for expressing the SO_2 absorption was developed with standard linear correlation techniques. Process parameters relevant to the type of absorption device were studied so that the SO_2 absorption efficiency could be estimated for similar operating systems.

Sulfur dioxide absorption efficiency for the cocurrent scrubber can be predicted by the following equation:

$$\begin{aligned} Y = 30.7 &+ 4.57(R) + 0.952(\Delta p) + 0.647(L/G) \\ &+ 15.16(I) - 2.751(I)^2 - 0.598(SL), \end{aligned}$$

when calcium oxide is used as absorbate. The scrubbing variables showing significant effect on the absorption (stoichiometry (R), throat pressure drop (Δp), liquid-to-gas ratio (L/G),

ionic strength (I), and slurry concentration (SL)) are conditions pertinent to most venturi-type scrubbers.

For sodium carbonate absorption, a less complicated correlation was developed, i.e.

$$Y = 25.36 + 3.105(\Delta p) - 0.0550(\Delta p)^2 + .0211(V)$$

Variables attributing liquid-phase resistance did not affect the SO_2 absorption efficiency (Y). Only throat pressure drop (Δp) and total gas flow (V) demonstrated significant sensitivity on efficiency.

The SO_2 absorption efficiency of the venturi was measured at 55% for maximum removal conditions using calcined limestone. Sodium carbonate allowed 80% removal at comparable conditions.

Venturi absorption with a mixture of sulfated lime and fly ash was also characterized as input to the imminent TVA/Shawnee scrubber demonstration project. Significantly lower absorption efficiencies were measured with sulfated lime/fly ash than with commercially calcined limestone.

Dolomitic lime (CaO·MgO) demonstrated excellent absorption efficiency in the single test made. The difference in absorption efficiency between calcium oxide, dolomitic lime, and sulfated lime/fly ash at comparable operating conditions are:

% Absorption Calcium Oxide	% Absorption Dolomitic Lime	% Absorption Sulfated Lime
32	64	16

The wetted film packed tower was studied with limestone (calcium carbonate) and correlations were developed for the SO_2 absorption (Y) only for the particular packing utilized. However, critical operating variables were identified. For example, sulfur dioxide removal was sensitive to inlet SO_2 concentration (ppm), limestone slurry concentration (% $CaCO_3$), and total slurry concentration (SL), as can be seen from the following:

$$Y = 165.05 - 0.0463(\text{ppm}) + 30.48\ (\%\ CaCO_3) - 9.126(\text{SL})$$

This correlation could predict the absorption efficiency to an accuracy of ± 1.9% for a limestone ground to 75%-200 mesh. A similar correlation was found for material containing 61%-200 mesh.

Long term scaling studies with $CaCO_3$ were not possible, but an 80-hour sustained operation was completed successfully with very favorable results. It is concluded that scaling can be controlled by direct limestone addition to the scrubbing circuit and that liquid-to-gas ratio and slurry concentration are primary variables.

It is recommended that additional on-site test work be conducted in the existing pilot unit employing limestone, sulfated lime/fly ash, and dolomitic limestone in order to determine the absorption efficiency and operational reliability with these materials.

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Experiments were carried out at the Tidd Power Station of Ohio Power, a subsidiary of American Electric Power. The cooperation of AEP and the Tidd Power Station personnel played a key part in executing this study.

Part of the work reported here was conceived and performed by the Tennessee Valley Authority personnel. The progress resulting from the limestone tests is due, in part, to TVA participation.

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I. INTRODUCTION

The ever-growing problem of sulfur dioxide atmospheric emission has been intensively studied in recent years by a host of researchers. Among the several processes that have been proposed, lime or limestone wet scrubbing holds the most promise for first generation SO_2 control systems. Simplicity in design, widespread abundance of limestone, and avoidance of by-product marketing complexities have all contributed to process acceptance economically and technically. While future generations of more economic SO_2 control systems might evolve, based upon a by-product recovery, legislative pressures demanding near-term flue gas desulfurization will require industries to make use of the best available, perhaps expedient, techniques.

Early in 1972, three prototype pilot plants will go on stream at TVA's Shawnee Steam Plant, Paducah, Kentucky, evaluating the Limestone Injection Wet-Scrubbing Process for sulfur dioxide and fly ash abatement. The pilot plant systems are being designed so that the key process variables affecting performance, process chemistry, scaleup, and economics will be defined. Each pilot plant will be equivalent to a 10-12 MW power generating station having a flue gas flow capacity of 30,000 cfm. The expected sulfur dioxide removal is 90% for a 4 percent sulfur fuel. Among the three parallel scrubbing trains, a venturi-type scrubber will be installed in series with an absorber. To facilitate the forthcoming prototype design and its operation, Cottrell Environmental Systems (CES) has carried out (under APCO Contract No. EHS-D-71-24) an experimental test program for sulfur dioxide removal in an existing two-stage pilot scrubber having a capacity of 1,000 cfm. The specific objectives of this study were:

1. Characterization of the maximum absorption capacity of a venturi scrubber,

2. Determination of the absorption capabilities of a venturi scrubber with calcium oxide injected into the boiler flue gas before the venturi (simulation of the Limestone Injection Wet-Scrubbing Process),

3. Comparison of the scrubbing characteristics of sulfated lime/fly ash material prepared at the Shawnee Steam Plant with commercially calcined limestone, and

4. Evaluation of other alkali material such as dolomitic lime, and limestone.

Other tasks were added to the program to study uncalcined limestone capabilities using the venturi in series with a packed tower contactor.

Initially, the venturi scrubber and the packed tower absorption performance were determined with sodium carbonate solutions. Results of the highly efficient sodium carbonate absorption were subsequently used as a guide in selecting the operating conditions for the calcium oxide tests.

The primary effort of the soda ash and calcium oxide experimentation and data analyses had been directed toward the understanding of the key variables affecting the SO_2 absorption within the venturi scrubber. However, where possible, the mass-transfer characteristics for the packed tower were also defined. Statistical analyses of the significant process variables have resulted in the development of simplified expressions for predicting SO_2 absorption. It is anticipated that these correlations for both the packed tower and the venturi scrubber can be used in estimating the absorption efficiencies for the Shawnee scrubber pilot program. At the same time the results of this work could guide the pilot plant design.

For limestone (calcium carbonate) wet scrubbing, the packed tower absorption capability and operating characteristics were the underlying objectives of the experimental program, while the venturi scrubber was considered secondary. Several limestone materials, comprising a range of chemical and physical properties, demonstrated high absorption efficiencies under properly controlled conditions. Tower scaling (encrustation buildup of reaction products on the packing) was studied carefully under various operating conditions. The limestone efficiency and scaling results were so encouraging that the original test program with calcium oxide and sodium carbonate was delayed and a new limestone test series was undertaken.

The second calcium carbonate test program explored further the scaling and absorption properties within the packed tower and the flooded disc scrubber. TVA and Radian Corporation have analyzed, on a limited basis, the slurry composition of the major process streams, and the results of their analysis have been invaluable in understanding important variables affecting the absorption as well as scaling.[1,2] Considerable effort has been applied in studying the slurry chemical composition and its relationship with absorption efficiency. Results of this limestone study have contributed significantly to the understanding of the hydrated lime and limestone absorption program. The limestone data as analyzed in this report will be useful in planning the Shawnee test work.

The process conditions for the limestone tests were based upon TVA bench scale experimental studies performed by the Chemical Development Division, Muscle Shoals, Alabama, and the Howden-ICI actual plant experience of the Fulham Power Station, London.[3] The results obtained with the limestone tests, while limited, are of great commercial significance in light of the Howden-ICI experience.

II. THEORY BACKGROUND

A. ABSORPTION RATE IN THE VENTURI

Using the concept of "transfer units" introduced by Chilton and Colburn,[4] the absorption efficiency for the venturi can be expressed as:

$$N_{oG} = -\ln (1-Y) \qquad \text{(II-1)}$$

where N_{oG} = number of overall gas phase transfer units,

Y = absorption efficiency, fraction.

When applying this expression, the product of the interfacial area per unit volume and the mass-transfer coefficients should be constant over the entire absorber. Also, an irreversible chemical reaction must be involved.

Although the interfacial area per unit volume for a Flooded Disc Scrubber (or Venturi) decreases down stream of the throat, Johnstone, et al[5] showed that a major portion of the mass-transfer takes place within a short distance from the interfacial generating point because of high droplet turbulence created as the liquid layers are atomized. If all of the mass-transfer takes place within a short distance down stream from the throat, where the interfacial area is relatively constant, then equation (II-1) is valid.

For the Venturi Scrubber, Galeano[6] found it convenient to express the number of transfer units in terms of gas flow and overall mass-transfer coefficients as in the following:

$$N_{oG} = Ka \frac{h M_c}{G} \qquad \text{(II-2)}$$

or

$$N_{oG} = \frac{KF}{Q} \qquad \text{(II-3)}$$

where K = overall mass transfer coefficient in terms of velocity, ft/hr.,

F = interfacial area, sq.ft.,

Q = gas flow rate, cu.ft./hr.,

a = absorbent surface per unit of volume absorber, sq.ft./cu.ft.,

h = height of absorber actively involved in the absorption, ft.,

M_c = molar density of gas, lb-mole/cu.ft.,

G = gas flow rate, lb-moles/(hr.)(sq.ft.).

Nukiyama and Tanasawa[7] studied the mean drop diameter produced by a gas-atomizing nozzle and developed an empirical relationship for the droplet diameter as a function of the gas velocity, liquid-to-gas ratio, surface tension, solution density, and liquid viscosity:

$$D_p = \frac{585}{V_t}\left(\frac{\sigma}{\rho}\right)^{\frac{1}{2}} + 597\left(\frac{\mu}{\sigma\rho}\right)^{.45}(L/G)^{1.5} \qquad \text{(II-4)}$$

where ρ, σ, and μ are the density, surface tension, and viscosity of the liquid, respectively.

D_p = mean droplet diameter, microns,

V_t = gas throat velocity, ft/sec.,

L/G = liquid-to-gas ratio, gal/1000 cu.ft.

Assuming the properties of air and water for the present study, this equation simplifies to:

$$Dp = 16050/V_t + 1.41\ (L/G)^{1.5} \qquad \text{(II-5)}$$

When the absorbent surface area per unit volume, a, is expressed in terms of D_p, the resulting equation is[7]:

$$a = \frac{L/G \times (30.5)^3 \times 10^{12}}{7.49 \times 1000 \times \frac{D_p^{\ 3}}{6}} \times \frac{D_p^{\ 2}}{(30.5)^2 \times 10^3} = 244\ \frac{L/G}{D_p} \qquad \text{(II-6)}$$

Combining equation (II-2), (II-6), and (II-5), the expression for the number of transfer units can be shown in terms of gas velocity, liquid-to-gas ratio, and active height of the absorber.

$$N_{oG} = \frac{244\ Kh\ L/G}{3600\ (16050 + 1.41\ L/G^{1.5} V_t)} \qquad \text{(II-7)}$$

The ratio (M_c/G) was replaced by $1/(3600\ V_t)$.

The active height for the venturi scrubber was studied by Johnstone, et al[5] and it was determined that the absorption rate is a maximum a short distance down stream from the liquid inlet where the relative velocity between the gas and liquid is the greatest. For a gas film controlled mass-transfer system, the absorption rate decreased to the same level as predicted for the quiescent drops within one foot after liquid injection. However, liquid film controlled absorption reached equilibrium within 2 to 3 inches after injection.

To compute the mass-transfer coefficients for Na_2CO_3 and CaO, the active height of the Flooded Disc Scrubber was set at 1 foot for each case throughout this report.

B. ABSORPTION RATE IN THE TOWER

Tower design is simplified by using the concept of transfer units which, for dilute solutions, are based on the definite integral

$$N_{oG} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} \qquad \text{(II-8)}$$

where y = concentration of solute in the bulk gas, mole fraction,

y* = concentration of solute in the gas in equilibrium with the bulk liquid, mole fraction.

The integral in equation (II-8) expresses the difficulty of a scrubbing solution to absorb solute from the gas. If Henry's law is applicable, equation (II-8) can be expressed as:

$$N_{oG} = \frac{\ln\left[(1-1/A)\left((y_1-mx_2)/(y_2-mx_2)\right) + 1/A\right]}{(1-1/A)} \qquad \text{(II-9)}$$

where A = absorption factor = $\frac{L}{mG}$, dimensionless

G = gas flow rate, lb-mole/(hr)(sq.ft.),

L = liquid flow rate, lb-mole/(hr)(sq.ft.),

m = slope of the equilibrium curve,

x = concentration in the bulk liquid, mole fraction.

If it can be assumed that the liquid is well mixed vertically, the chemical reaction is irreversible, and the product of the transfer coefficient per unit area and the interfacial area per unit of liquid volume is constant along the vertical path of integration, then equation (II-9) can be expressed as follows:

$$N_{oG} = -\ln (1-Y) \qquad \text{(II-10)}$$

Chilton and Colburn[3] have developed a relationship for the number of transfer units for a packed tower

$$N_{oG} = \frac{\bar{K}\, a\, P\, Z}{G} \qquad \text{(II-11)}$$

where $\bar{K}$ = overall mass transfer coefficient, lb-mole/(hr)(sq.ft.)(atm),
a = surface area per unit volume of packing, sq.ft./cu.ft.,
P = total pressure of the system, atm.,
Z = height of tower, ft.,
G = gas flow rate, lb-mole/(hr)(sq.ft.),
N_{oG} = number of transfer units as defined by equation (II-10).

In terms of locally applicable coefficients, the rate of mass-transfer is given by

$$F_{SO_2} = k_g\,(y_{SO_{2g}} - y_{SO_{2i}}) = k_\ell\,(X_{SO_{2i}} - X_{SO_{2\ell}}) \qquad \text{(II-12)}$$

where F_{SO_2} = mass-transfer flux, lb-mole/(hr)(sq.ft.),

k_ℓ = liquid phase mass-transfer coeffieients, lb-mole/(hr)(sq.ft.)(mole fraction),

k_g = gas phase mass-transfer coefficient, lb-mole/(hr)(sq.ft.)(mole fraction),

y_{SO_2} = concentration of SO_2 in the gas, mole fraction,

x_{SO_2} = concentration of SO_2 in the liquid, mole fraction,

i = refers to the interface,

g = refers to the bulk gas stream,

ℓ = refers to the bulk liquid stream.

Since $y_{SO_{2i}}$ and $x_{SO_{2i}}$ are extremely difficult to measure, overall mass-transfer coefficients are employed to eliminate the dependence on the interfacial compositions.

$$F_A = K_{oG} \; y_{SO_{2g}} - y_{SO_2}{}^*) = K_{oL} \; (x_{SO_2}{}^* - x_{SO_{2\ell}}) \qquad \text{(II-13)}$$

where K_{oG} = overall gas phase mass-transfer coefficients, lb-mole/(hr)(sq.ft.)(mole fraction),

K_{oL} = overall liquid phase mass-transfer coefficient, lb-mole/(hr)(sq.ft.)(mole fraction),

$y_{SO_2}{}^*$ = concentration of SO_2 in the gas in equilibrium with the bulk concentration in the liquid mole fraction,

$x_{SO_2}{}^*$ = concentration of SO_2 in the liquid in equilibrium with the bulk concentration in the gas, mole fraction.

For irreversible chemical reaction in the packed tower, $y_{SO_2}{}^{*} = 0$. Therefore

$$F_A = K_{oG}\, y_{SO_{2g}} \qquad \text{(II-14)}$$

As can be seen from equation (II-14), the rate of mass-transfer is proportional to the mole fraction of SO_2 in the gas stream.

In terms of the two-resistance theory with a solute exhibiting a partial pressure in accordance with Henry's law

$$\frac{1}{K_{oG}} = \frac{1}{k_g} + \frac{m}{k_\ell} \qquad \text{(II-15)}$$

where m = slope of the equilibrium curve.

The rapid, irreversible reaction which occurs in a tower also gives rise to an equilibrium curve with a slope of nearly zero for the SO_2 concentration under consideration. Equation (II-15) reduces to

$$K_{oG} = k_g \qquad \text{(II-16)}$$

thus placing all the resistance to mass-transfer in the gas phase.

Although the major resistance derived in equation (II-16) is in the gas phase, the lack of an equilibrium condition does not necessarily eliminate the possibility of a liquid phase resistance. Other factors must be considered, namely, the diffusivity of the solute in the liquid-phase, the concentration of the unreacted reagent, the rate of diffusion of the reagent, the rate of dissolution for heterogenous slurries, etc.

Using the two-resistance theory as a basis, it would be better to express the overall transfer coefficient by the following:

$$\frac{1}{K_{oG}} = \frac{1}{k_g} + \frac{1}{k_\ell} \qquad \text{(II-17)}$$

Here the type of liquid phase resistance is not defined, in terms of vapor/liquid equilibrium, and in fact can incorporate all liquid phase resistances.

C. PROCESS CHEMISTRY

The removal of SO_2 from gas streams by absorption is well known. Water itself is a relatively poor solvent for SO_2; consequently, its use would entail vast quantities of water to reduce SO_2 levels appreciably. Further, the SO_2 is readily released by environmental influences (temperature, pH, etc.).

The use of alkaline solutions to fix SO_2 is frequently practiced. Thus, SO_2 reacts as follows with soluble hydroxide.

$$SO_2\ (g) \rightleftarrows SO_2\ (aq)$$

$$SO_2(aq) + H_2O \rightleftarrows H^+ + HSO_3^-$$

$$HSO_3^- + OH^- \rightarrow SO_3^= + H_2O$$

Thus, the absorption proceeds via a $SO_3^=$ rich liquor.

The partial pressure of SO_2 in this alkaline system is essentially proportional to the square of the hydrogen ion concentration. In $SO_3^{=}$ solutions, this concentration is quite small and the corresponding SO_2 pressure is negligible. In fact, at ordinary temperatures, a level of $SO_2/Na \cong 0.9$ can be achieved before SO_2 back pressure becomes significant.

The use of lime for absorption of SO_2 from sulfuric acid tail gas, in the absence of CO_2 proceeds:

$$Ca(OH)_2 + SO_2 + H_2O \dashrightarrow CaSO_3.2H_2O \downarrow$$

In the presence of CO_2, the situation is more complex. Pearson,et al[8] proposed possible lime and limestone reactions for combustion gas.

$$CaO + H_2O \dashrightarrow Ca(OH)_2$$

$$Ca(OH)_2 + CO_2 \dashrightarrow CaCO_3 + H_2O$$

$$CaCO_3 + CO_2 + H_2O \dashrightarrow Ca(HCO_3)_2$$

$$Ca(HCO_3)_2 + SO_2 + H_2O \dashrightarrow CaSO_3.2H_2O \downarrow + 2CO_2$$

$$CaSO_3.2H_2O + 1/2O_2 \dashrightarrow CaSO_4.2H_2O \downarrow$$

While it is possible to define the equilibrium situation based on a knowledge of the components present and the final equilibrium conditions (temperature, etc.), the kinetic competition of various important reactions is not well defined.

Both carbon dioxide and sulfur dioxide are weak acids, the former being weaker. The presence of carbon dioxide alters appreciably the kinetics of SO_2 absorption. It is expected that absorption will proceed through a calcium carbonate step as a result of the approximate 50+ ratio of CO_2/SO_2 in the gas.

In the slurry at pH ∿ 5-6, sufficient levels of dissolved carbonate salts exist to react with the HSO_3^-. Calcium sulfite precipitation results.

Oxidation of sulfite species is also important, resulting ultimately in calcium sulfate, a sparingly soluble species. Conversion of $SO_3^=$ to $SO_4^=$ influences the equilibrium partial pressure of SO_2 over the liquor.

III. PROCESS EQUIPMENT

A. PILOT PLANT LAYOUT

The two-stage absorption operation used in this work was designed for a test program that would allow maximum flexibility and versatility.

A layout of the pilot plant system is illustrated in Figure III-1. The complex piping network shown was necessary to accommodate the 13 operating modes planned. Slurry flow rates to the scrubber and other process units were measured by venturi-type flow meters. In-line pH probes (Leeds-Northrup) were installed at the discharge of each scrubber. Immersion-type pH elements were placed in the clarifier and the hold tank.

All temperature measurements were made with iron-constantan thermocouples except the wet bulb temperature on the inlet.

The absorption section of the pilot plant contained the Flooded Disc Scrubber (FDS) in series with a packed tower, as illustrated in Figure III-2. The flue gas, containing both particulates and sulfur dioxide, passed first through the FDS where the entering gas quickly cooled to its dew point (120°F). Scrubbing solution or slurry entering the absorber tangentially above the throat flowed cocurrently through the FDS and into a cyclonic demister. The fly-ash-stripped gas then passed vertically through a conical hat gas/liquid splitter before entering the packed tower.

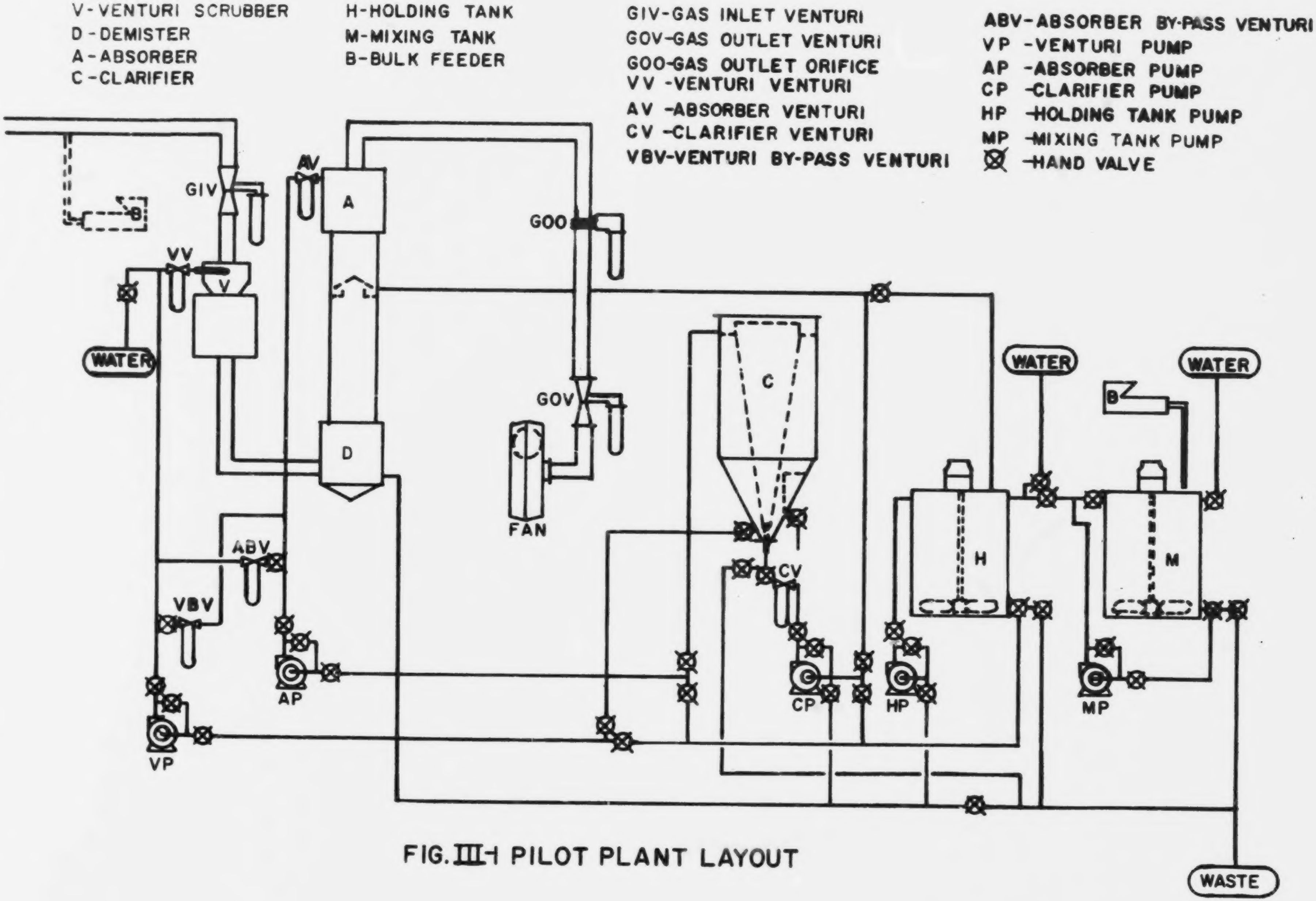

V-VENTURI SCRUBBER
D-DEMISTER
A-ABSORBER
C-CLARIFIER
H-HOLDING TANK
M-MIXING TANK
B-BULK FEEDER
GIV-GAS INLET VENTURI
GOV-GAS OUTLET VENTURI
GOO-GAS OUTLET ORIFICE
VV-VENTURI VENTURI
AV-ABSORBER VENTURI
CV-CLARIFIER VENTURI
VBV-VENTURI BY-PASS VENTURI
ABV-ABSORBER BY-PASS VENTURI
VP-VENTURI PUMP
AP-ABSORBER PUMP
CP-CLARIFIER PUMP
HP-HOLDING TANK PUMP
MP-MIXING TANK PUMP
⊗-HAND VALVE
GIV
B
VV
V
WATER
AV
A
D
GOO
GOV
FAN
ABV
VBV
AP
VP
C
CV
CP
H
HP
WATER
B
WATER
M
MP
WASTE

FIG. III-1 PILOT PLANT LAYOUT

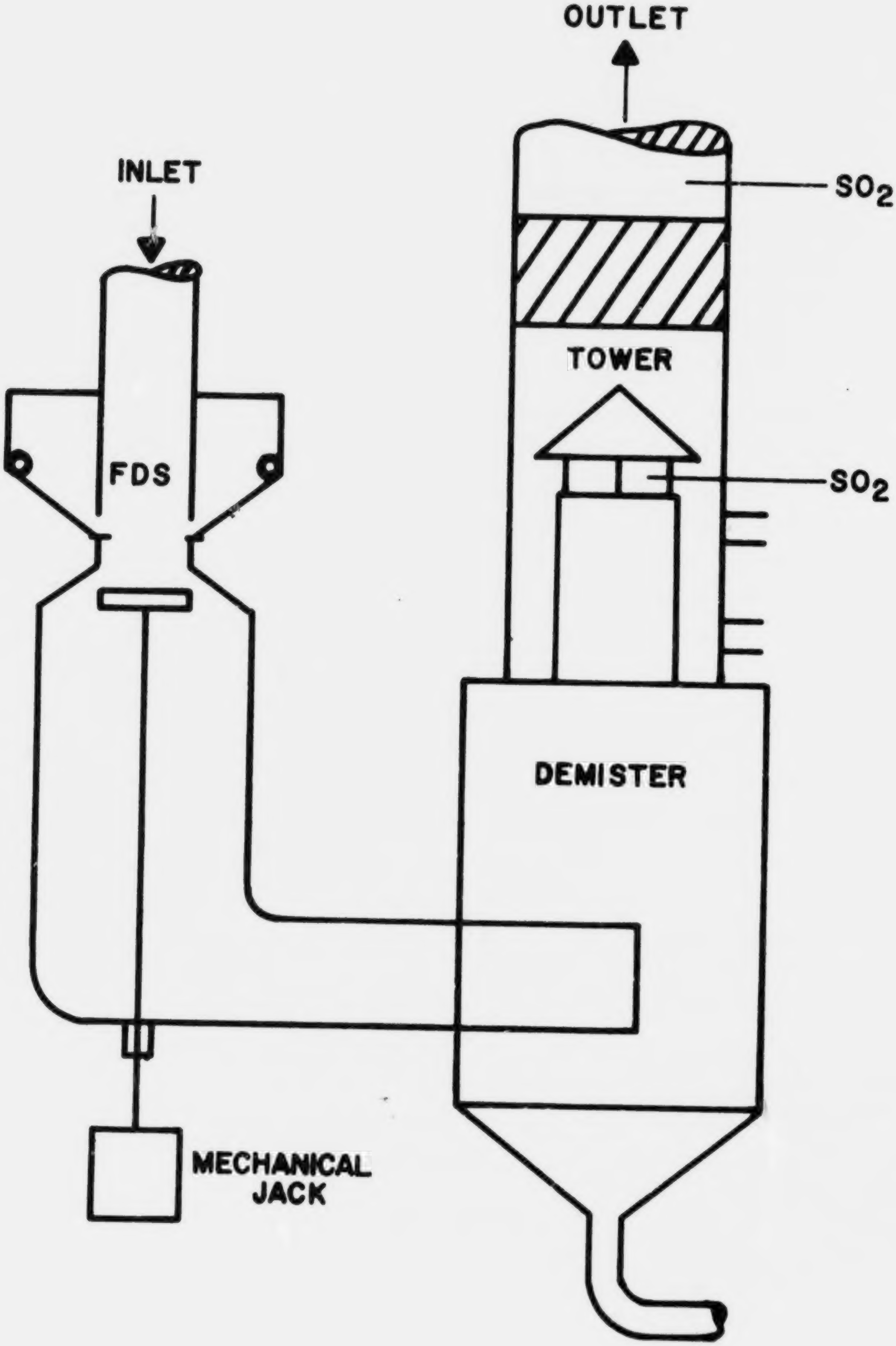

OUTLET
INLET
SO_2
TOWER
FDS
SO_2
DEMISTER
MECHANICAL JACK

FIG. III-2 PILOT PLANT SCHEMATIC DIAGRAM

The FDS scrubber is a cocurrent absorber with a variable throat orifice. Pressure drop across the scrubber is varied by adjusting the disc position within a venturi throat, Figure III-3.

The packed tower is a countercurrent absorption device containing a packing with low pressure drop characteristics and high specific surface (68 sq.ft./cu.ft.). The packing section was fabricated from rigid corrugated sheets of asbestos coated with neoprene. It was five feet in height and sixteen inches in diameter. Pressure drop across the packing at gas velocities between 8 and 10 feet per second under well irrigated conditions is approximately one inch of water.

It will become evident throughout the body of this report that many modes of operation were tested over a wide range of conditions. Tank sizes and clarifier volume listed below were sized and selected on the basis of studying the process variables.

Process Units	Size	Material Of Construction
FDS Scrubber	6" to 8" diameter See Figure III-3	316SS
Packed Tower	16" diameter x 5'	316SS
Tower Slurry Tank	1500 gallons agitated	CS
Mixing Tank	1500 gallons agitated	CS
FDS Slurry Tank	55 gallons	CS
Cyclonic Demister	4' diameter x 4'	SS
Clarifier	1200 gallons	CS
Blower	1000 cfm at $\Delta p = 40$ inches H_2O	CS

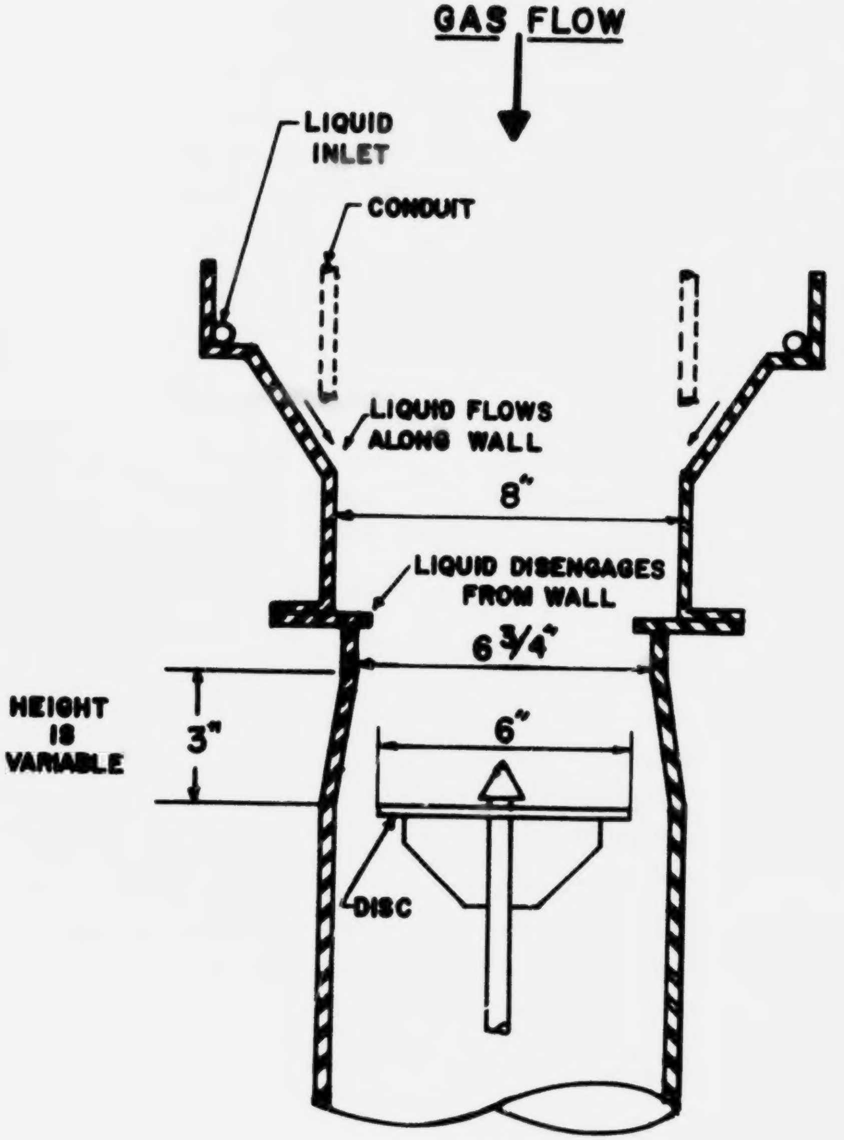

GAS FLOW
LIQUID INLET
CONDUIT
LIQUID FLOWS ALONG WALL
8"
LIQUID DISENGAGES FROM WALL
6 3/4"
HEIGHT IS VARIABLE
3"
6"
DISC

FIG. III-3 DIMENSIONS FOR FLOODED DISC SCRUBBER

B. ANALYTICAL

Sulfur dioxide concentration in the gas phase was determined by an Enviro-Metrics SO_2 analyzer Model S-64S and by titrimetric techniques. Gas samples were drawn from the scrubbers, filtered and pumped through the instrument as shown schematically in Figure III-4. Each analyzer was calibrated daily by use of standard gas obtained from compressed gas cylinders. Throughout any test period, the instruments were continually checked for proper calibration.

Samples were drawn from the flue gas by a Gast pump rated at 3 cfm. The sample gas passed through a fiber glass filter before entering the pump. Most of the particulates were removed in the filter. A high gas flow of 1 to 1.5 cfm passed through the coarse rotameter and into the venting lines. A small sample bypass stream from the inlet to the coarse rotameter passed through the instrument. A maximum rate of 25,000 cc per hour flowed through the analyzers.

Instrument sampling and calibration were timed and controlled so that the same volume of gas passed through the instrumentation. The analyzer meter reading, indicating the SO_2 level, was recorded and then the SO_2 concentration was calculated for each test.

Instrument response was checked by introducing calibration gases at two levels of SO_2 concentration. A linear response was measured for instrument readings between 30 and 90% on the meter scale. Wet analysis of the SO_2 concentration confirmed the results of the instrumentation throughout the test program. Sulfur dioxide concentration ranged from 950 to

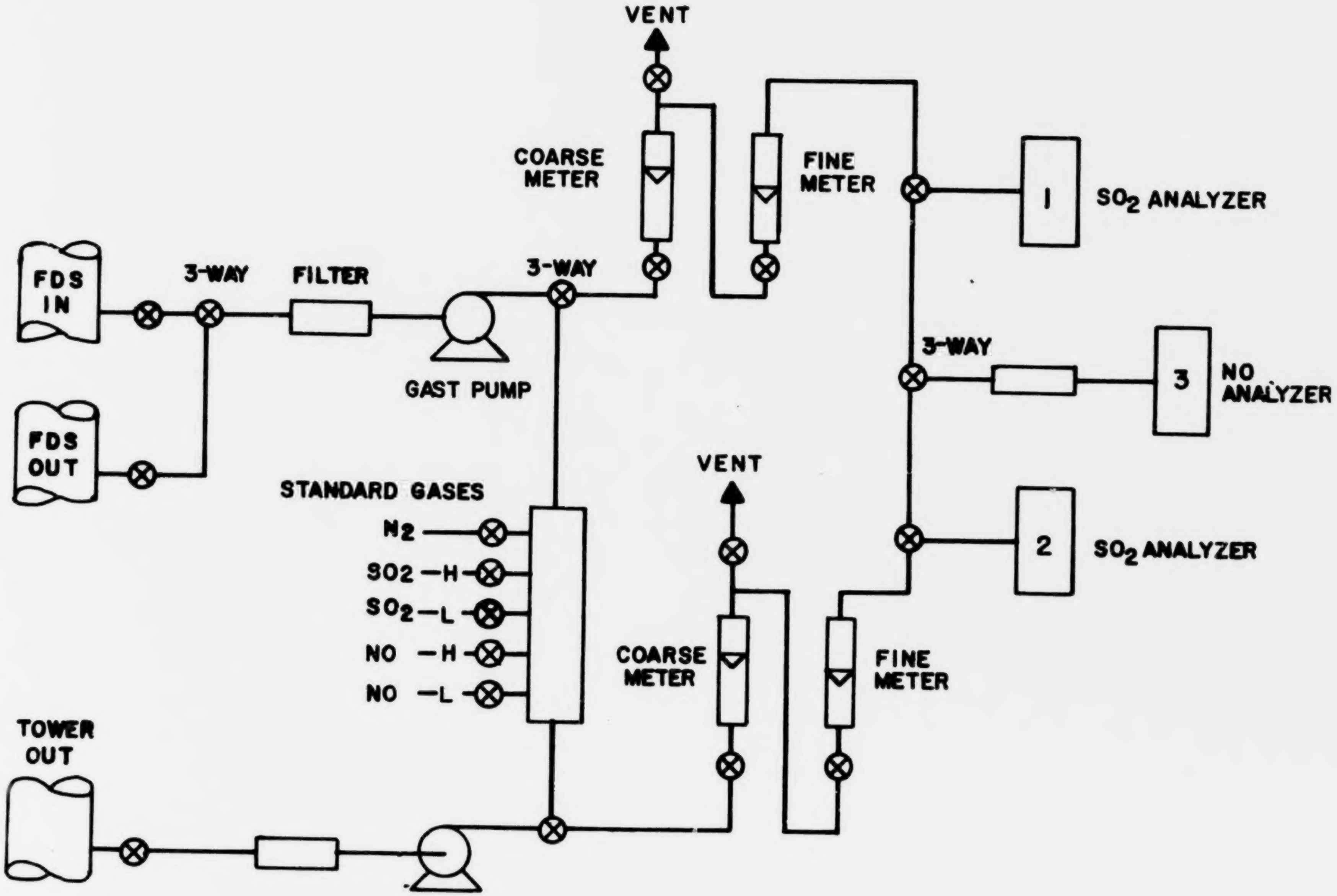

VENT
COARSE METER
FINE METER
SO$_2$ ANALYZER
1
FDS IN
3-WAY
FILTER
3-WAY
GAST PUMP
3-WAY
3
NO ANALYZER
FDS OUT
VENT
STANDARD GASES
N2
SO2 —H
SO$_2$—L
NO —H
NO —L
2
SO$_2$ ANALYZER
COARSE METER
FINE METER
TOWER OUT

FIG. III-4 SCHEMATIC DIAGRAM FOR SO$_2$ ANALYTICAL SYSTEM

2350 ppm. Fly ash concentrations of the inlet gas were not measured during this program; however, previous studies measured the particulate at 2.4 to 3.5 grains/SCFD.

Nitric oxide analysis was attempted early in the experimental program. The gas sample was passed through the same sampling equipment that was used for the SO_2 analysis and into the NO instrument. A three-way valve was used to divert the gas sample from the SO_2 analyzer to the NO instrument. A fixed bed of Mallcosorb removed the SO_2 from the gas stream before it entered the NO unit.

Results of the nitric oxide analysis were erratic from the beginning. The outlet gas concentration on the packed tower was sometimes higher than that measured at the inlet to the FDS. To avoid undue time losses in the test program, the NO analyses, which were a secondary part of the planned tasks, were abandoned.

1. Chemical Reagents

Calcium oxide used in all experimental work calling for calcined limestone was donated to the project by Basic Chemicals, Cleveland, Ohio, Table A-1 of Appendix A.

Dolimitic lime used in comparison to the calcium oxide was 99% MgO.CaO with approximately 50% - 200 mesh. J. E. Baker Company, York, Pennsylvania, donated this material to the project.

Limestone and lime reagent used for the test program studying the packed tower characteristics was provided by TVA. A list of the chemical compositions of these materials are shown in Table B-1, Appendix B.

IV. RESULTS AND DISCUSSION

The experimental test plan for the sodium carbonate and the calcined limestone required thirteen operating modes; each studying a specific variable or condition. A description for each of the planned tasks is given in Table A-2 of Appendix A. Throughout the following section, reference will be made to these tasks as the results are discussed.

A. SODIUM CARBONATE

Absorption efficiency experiments with sodium carbonate were made to determine the maximum mass-transfer properties of the FDS scrubber and the packed tower. Gas flow through the absorbers was adjusted between 300 to 900 ACFM* while the liquid-to-gas ratios were varied over the range of 5 to 14 gallons per 1000 cf for the FDS, and 7 to 34 gallons per 1000 cf for the tower. Operating modes for this test series are illustrated in Figures IV-1 and IV-2.

Sodium carbonate and water were mixed continuously in an agitated tank from which a carbonate solution was withdrawn and pumped to the disc scrubber and/or the tower. Stoichiometry to each absorber was controlled by adjusting the carbonate feed rate and the liquid flow rate.

Sodium carbonate stoichiometry for the inlet SO_2 varied between 90 and 250%.

---

* Flow rate base on outlet conditions at the dewpoint temperature.

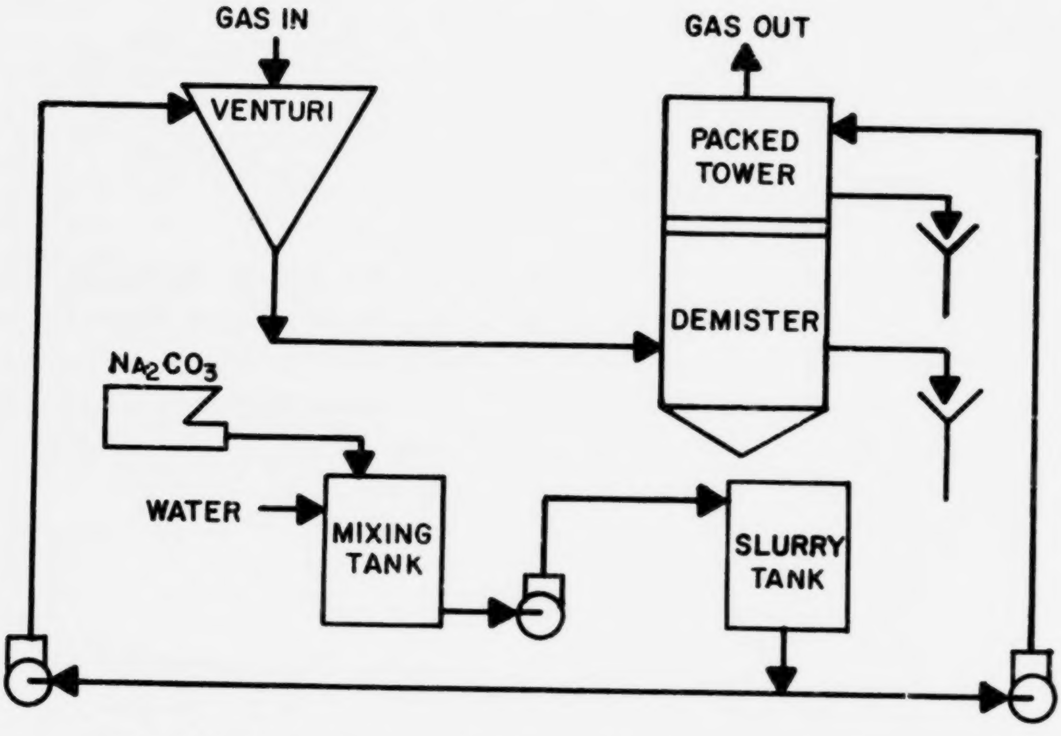

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
NA_2CO_3
WATER
MIXING TANK
SLURRY TANK

FIG. IV-1 TWO STAGE SODIUM CARBONATE SCRUBBER-TASK II A&B

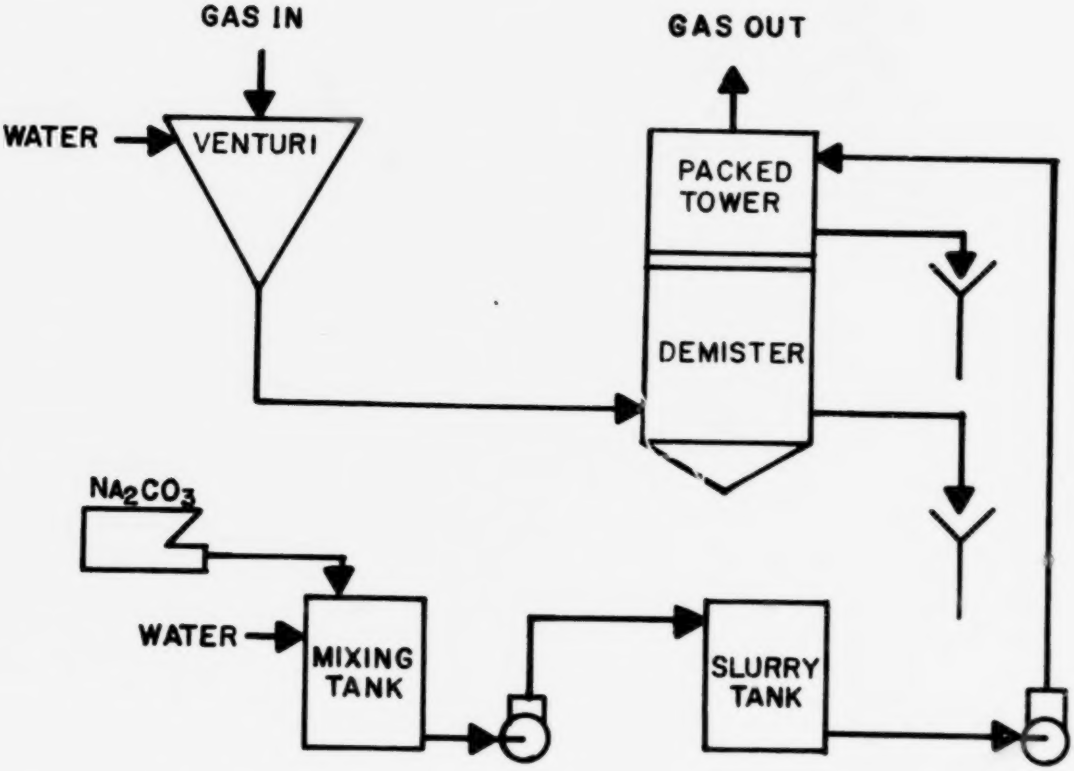

GAS IN
GAS OUT
WATER
VENTURI
PACKED TOWER
DEMISTER
NA_2CO_3
WATER
MIXING TANK
SLURRY TANK

FIG. IV-2 TOWER ABSORPTION TESTS FOR SODIUM CARBONATE-TASK IIC

1. FDS Results - Na_2CO_3

Absorption efficiency characteristics for the FDS were measured for five levels of gas flow, five levels of disc pressure drop, four levels of liquid-to-gas ratio, and the above range of stoichiometries.

SO_2 absorption efficiencies increased rapidly from 30 to 73% as the pressure drop across the scrubber increased from 1 to 13 inches of water. At high pressure differentials, 13 to 25 inches of water, the SO_2 absorbed increased more slowly and reached 80 to 83%. Stoichiometry did not significantly affect the efficiency. However, slightly greater absorption occurred at increased gas flow rates. A linear regression analysis for 46 selected run conditions resulted in the following expression:

$$Y = 25.36 + 3.105\,(\Delta p) - 0.0550\,(\Delta p)^2 + .0211(V) \qquad \text{(IV-1)}$$

Y = SO_2 absorption efficiency, %,
where Δp = pressure drop across the FDS, inches of H_2O,
V = volume of gas flow, CFM at 110°F. and approximately 390 inches of H_2O.

The positive coefficient for the gas flow does not follow any expected absorption mechanism. Increasing gas velocity at the entry of the scrubber may have caused significant shearing action on the liquid as it disengaged from the wall, hence some interfacial surface area generation and mass-transfer could have occurred before the throat. Similarly increased gas velocity downstream from the throat can contribute to mass-transfer. The statistical results of the carbonate regression analysis and the limitations for equation (IV-1) are listed in Table IV-1. The experimental data used in the regression analysis are listed in Table A-3, Appendix A.

With the absorption efficiency depending almost exclusively on the pressure drop and showing independence of stoichiometry and liquid-to-gas ratio, a gas film controlled mass-transfer appears to exist. Using equation (II-7), the overall mass-transfer coefficients for each data set were calculated. These results are illustrated in Figure IV-3.

A regression analysis on the overall mass-transfer coefficients yielded the following expression:

$$\bar{K}_G = 4.16 + 0.53\ (V_t) - 1.73\ (L/G) \qquad \text{(IV-2)}$$

where L/G = liquid-to-gas ratio, gal. per 1,000 cf.,
V_t = throat velocity, ft./sec.,
$\bar{K}_G$ = overall mass-transfer coefficient, lb-mole/(hr)(sq.ft.)(atm).

Statistical limitations for equation (IV-2) are given in Table IV-2.

2. Packed Tower Results - Na_2CO_3

An average of 95% SO_2 absorption efficiency was obtained when the tower operated at a gas velocity of 7.7 to 13 feet per second and liquid flows of 6 to 34 gallons per 1000 cf. Sodium carbonate passed through the tower and drained into the pond as shown in Figure IV-2.

TABLE IV-1

STATISTICAL PARAMETERS FOR THE SODIUM CARBONATE CORRELATION FOR FDS*

Number of Data Points = 46, See Table A-3

Correlation Coefficient = 0.8890

Standard Error for the Estimate = 7.26%

Significance of Regression (F) = 52.8

Pressure Drop Range (Δp) = 1 to 25 inches of H_2O

Gas Velocity Range (throat) = 50 to 254 ft./sec.

Gas Volume Range (V) = 300 to 920 CFM at 110°F. and 380 inches of H_2O.

Stoichiometric Range = 0.9 to 2.5

Sulfur Dioxide Inlet Range = 1000 to 2350 ppm.

---

* $Y = 25.36 + 3.105 (\Delta p) - 0.0550 (\Delta p)^2 + .0211(V)$

FIGURE IV-3

MASS TRANSFER COEFFICIENTS FOR SODIUM CARBONATE
THROAT VELOCITY BETWEEN 50 AND 250 FEET PER SECOND

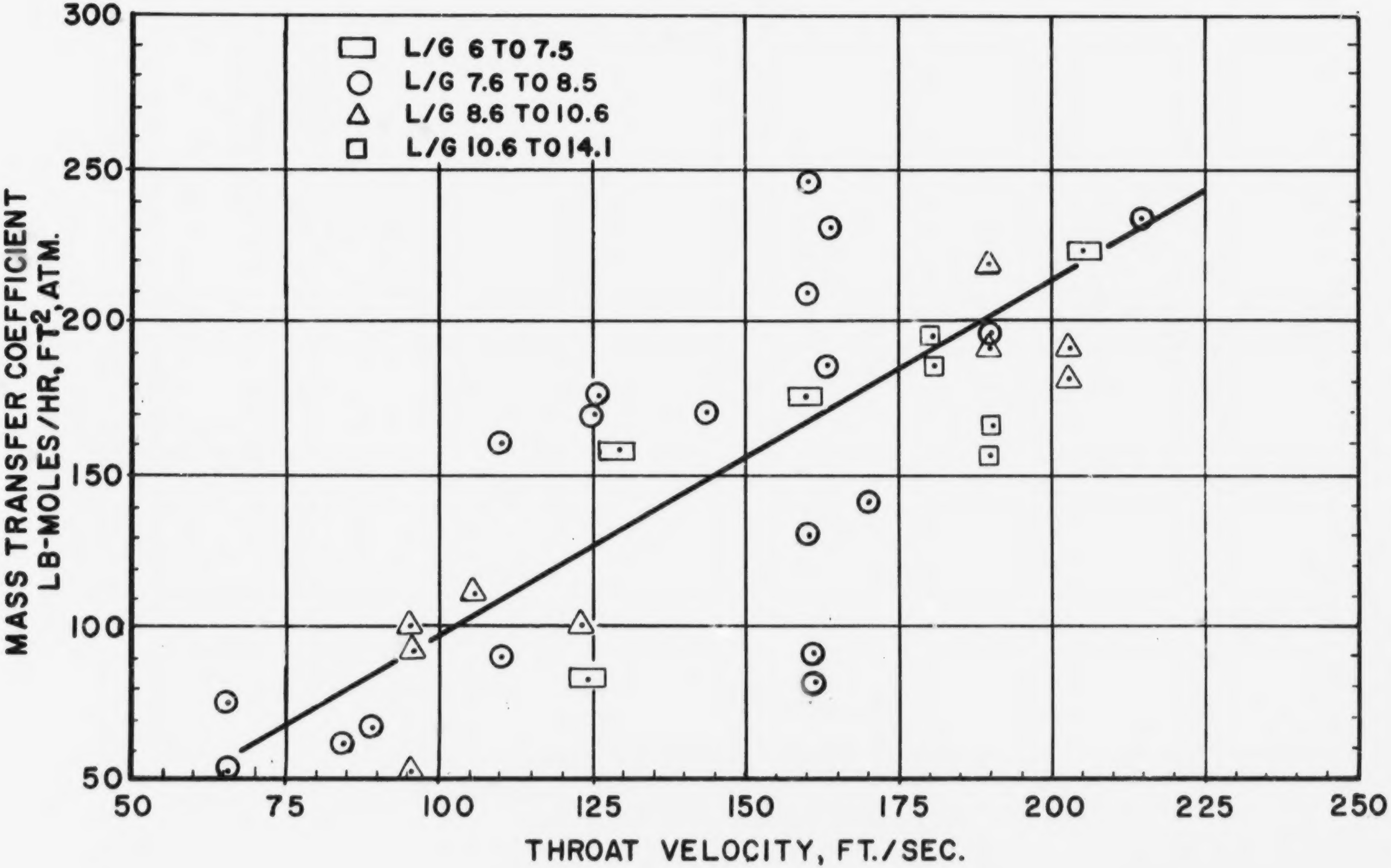


TABLE IV-2

STATISTICAL PARAMETERS FOR SODIUM CARBONATE MASS-TRANSFER CORRELATION

Number of Data Points = 46, See Table A-3

Correlation Coefficient = 0.864

Standard Error for Y-Data = 29.3

Standard Error for Estimate = 15.1

Significance of Regression (F) = 63.6

Mass-Transfer Coefficient Range (K_G) = 22.2 to 136.0 lb-moles/(sq.ft.)(hr)(atm).

Throat Velocity Range (V_t) = 51 to 254 ft/sec.

Liquid-to-Gas Ratio Range (L/G) = 6.7 to 14.1

Sulfur Dioxide Inlet Concentration Range = 1,000 to 2,350 ppm.

Overall mass-transfer coefficients for the tower were computed from equation (II-11) and ranged from 0.6 to 1.7 (lb-moles)/(hr.)(sq.ft.)(atm), see Table IV-3. A regression analysis of the overall mass-transfer mechanism in terms of gas flow and liquid rate was made using the following equation:

$$K_{oG} = C\,\bar{G}^{a},\,\bar{L}^{b} \qquad \text{(IV-3)}$$

The statistical analysis indicated a strong dependence of the overall mass-transfer coefficient on the gas rate and a negligible effect of liquid rate, as indicated by the following equation:

$$K_{oG} = 6.69(10)^{-5}\,\bar{G}^{1.17}\,\bar{L}^{0.026} \qquad \text{(IV-4)}$$

where K_{oG} = lb-mole /(hr)(sq.ft.)(atm),

$\bar{G}$ = gas mass velocity, lb/(hr)(sq.ft.),

$\bar{L}$ = liquid mass velocity, lb/(hr)(sq.ft.).

The mass-transfer coefficients for sodium carbonate are compared in Figure IV-4 with sodium hydroxide experiments performed with the same type of packing in the laboratory. The slightly lower absorption efficiency of the sodium carbonate indicates some liquid film resistance.

B. FDS RESULTS - CALCIUM OXIDE

1. Open-Loop: Dry Injection and Wet Slurry - Task III and IV

Several operational modes were selected for the calcium oxide program so that the variables influencing performance could be isolated and their effects measured.

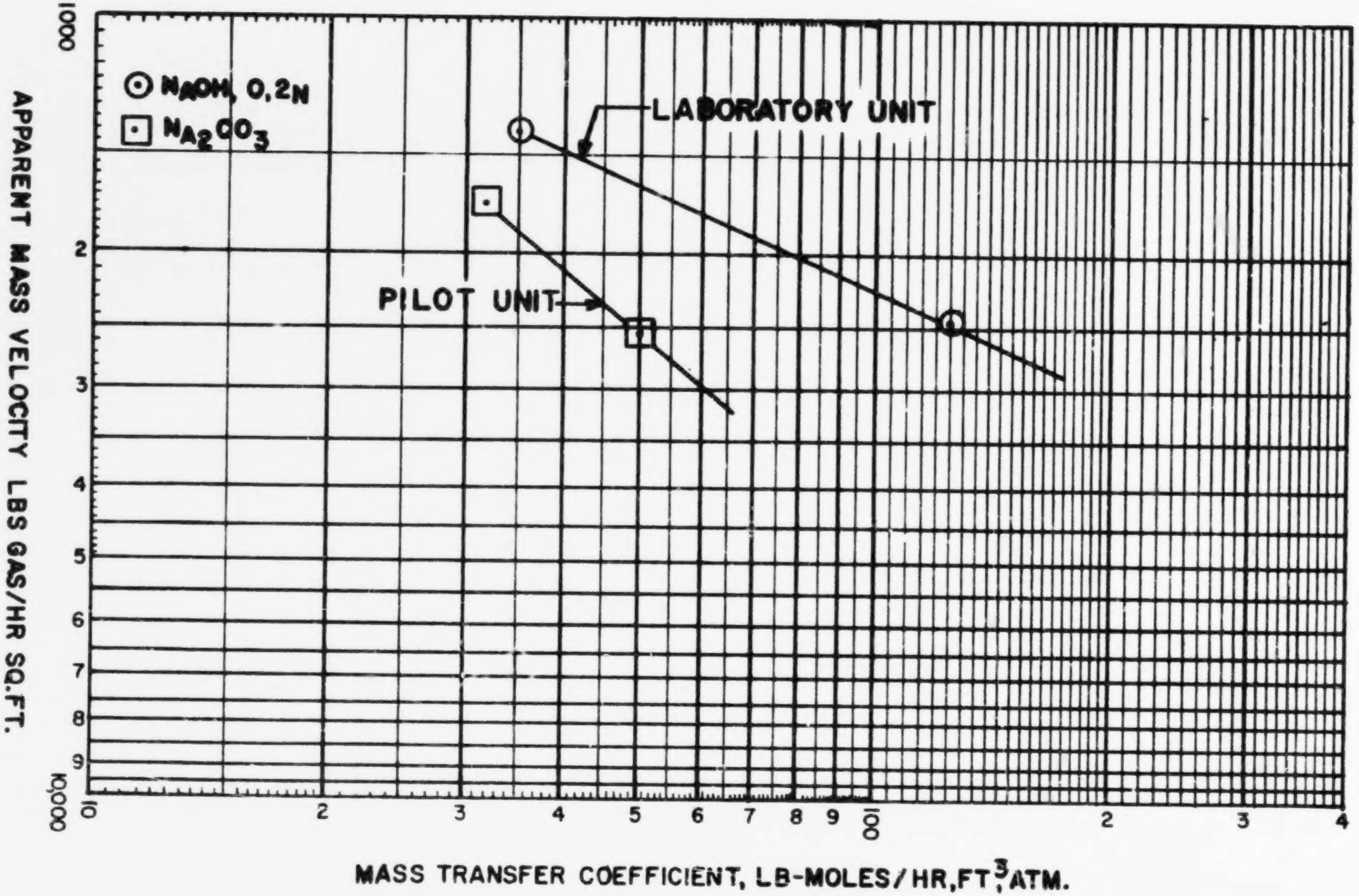
TABLE IV-3

MASS-TRANSFER COEFFICIENTS FOR SODIUM CARBONATE IN WETTED FILM PACKED TOWER

Inlet SO_2 Conc. (PPM)	Outlet SO_2 Conc. (PPM)	Gas Velocity (Ft/Sec)	Liquid & Gas Ratio (Gals/1000CF)	Mass Transfer Coefficient $\frac{\text{(lb-moles)}}{\text{(hr.)(sq. ft.)(atm.)}}$
860	16	7.8	10	0.731
680	8	7.8	11.9	0.816
700	4	7.8	15.8	0.949
820	15	7.8	22.4	0.733
730	30	7.8	6	0.586
830	35	7.8	6	0.583
800	18	7.7	30	0.71
960	26	7.8	30	0.66
930	35	7.8	34	0.60
780	28	7.8	34	0.609
910	20	12.6	6	1.137
878	7	12.6	6	1.419
1000	20	12.6	7.5	1.151
852	6	12.6	18.7	1.465
852	10	12.6	18.8	1.327
780	20	12.6	7.4	1.089
750	17	12.6	9.8	1.125
900	20	12.6	12.2	1.133
804	11	13	11.9	1.311
1000	30	12.6	12.2	1.043
960	4	12.6	12.2	1.668
1000	38	12.6	18.8	0.999
1040	32	12.6	18.8	1.032
860	16	12.8	18.5	1.2
700	11	12.6	.21	1.232

FIG. IV-4

MASS TRANSFER COEFFICIENT AS A FUNCTION OF GAS MASS VELOCITY



Initially, an open-loop dry injection operation was employed with the venturi scrubber as shown in Figure IV-5. The process conditions were fixed at a liquid-to-gas ratio of 10 gallons per 1,000 cf, and a pressure drop across the venturi of 10 inches W.G. Stoichiometry varied between 0.8 to 2.1 moles of CaO/mole of SO_2.

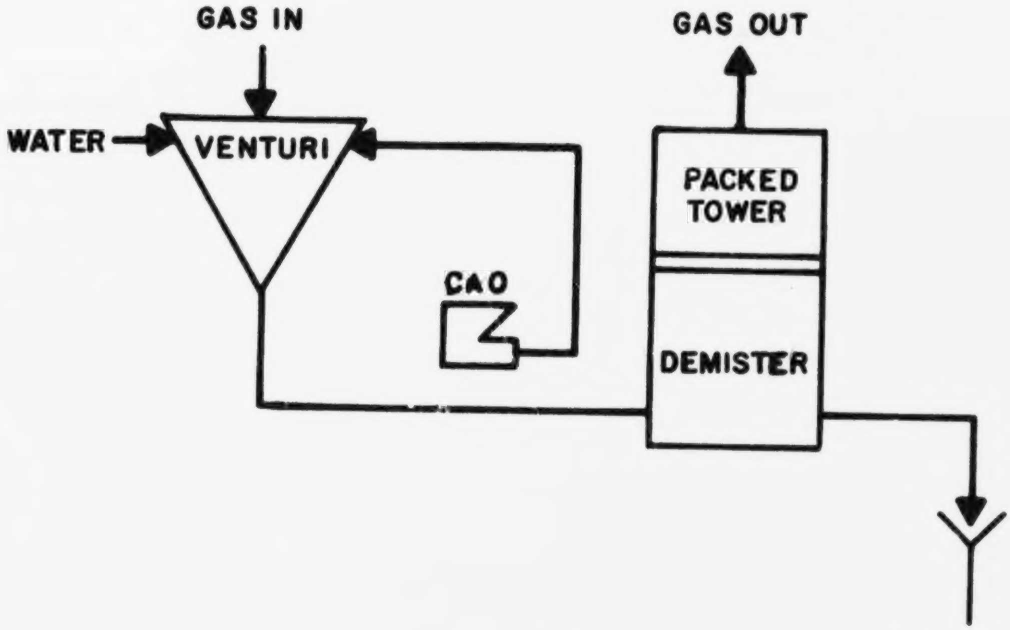

GAS IN
GAS OUT
WATER
VENTURI
PACKED TOWER
CAO
DEMISTER

FIG. IV-5 DRY INJECTION WITHOUT RECIRCULATION TASK III

Absorption efficiency improved with increasing lime input. At stoichiometric lime conditions, approximately 30% removal was achieved in the venturi while at twice the equivalent calcium oxide, absorption reached 45% efficiency. These results are summarized in Figure IV-6.

SO_2 absorption efficiency for wet slurry feed exhibited higher removal than the dry injection. Using the operating mode shown in Figure IV-7, with a liquid-to-gas ratio (FDS) of 10 gallons per 1,000 cf and pressure drop across the FDS of 10 in. W.G., 40 to 50% SO_2 absorption was achieved, i.e. between 10 to 15% more SO_2 was removed by the wet slurry than by the dry injection, again see Figure IV-6.

Absorption measurements made with water containing no alkali ranged between 2 and 4% at the same operating conditions as illustrated in Figure IV-6. Hence, the efficiency difference between dry injection and wet slurry can be contributed to the method for lime addition, i.e. water absorption for the dry injection could contribute only a small fraction to the total removal.

Detailed operating data for these experiments are presented in Appendix A, Tables A-4 and A-5.

2. Closed-Loop: Dry Injection/Wet Slurry Combination

Due to practical, as well as economical, considerations, most commercial wet scrubbing processes will be based on a closed-loop system. For the first operating mode using complete solution recirculation, a lime slurry was pumped to the venturi while a clarified solution was passed through the tower, as shown in Figure IV-8.

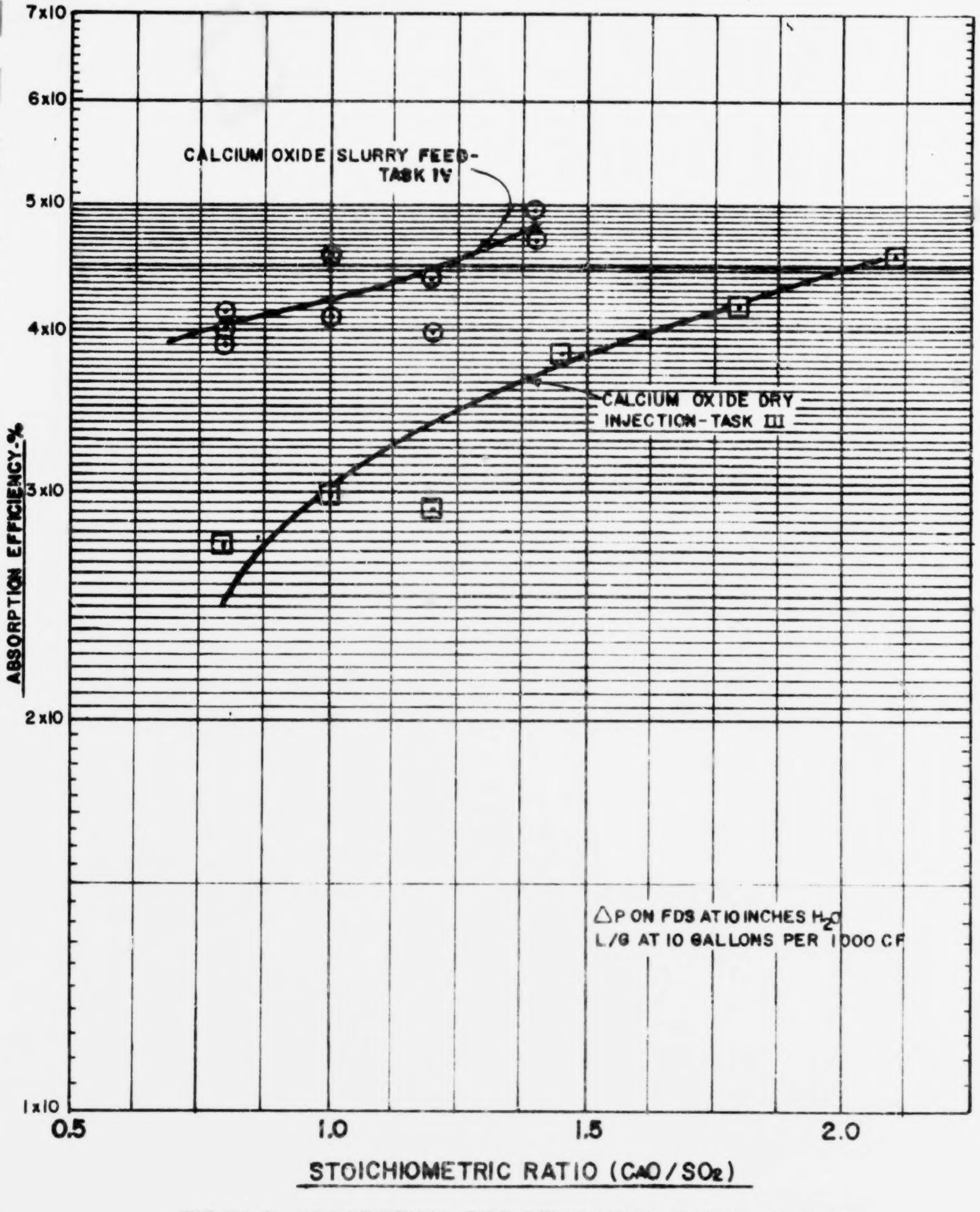


FIG. IV-6 ABSORPTION EFFICIENCY FOR TASKS III & IV

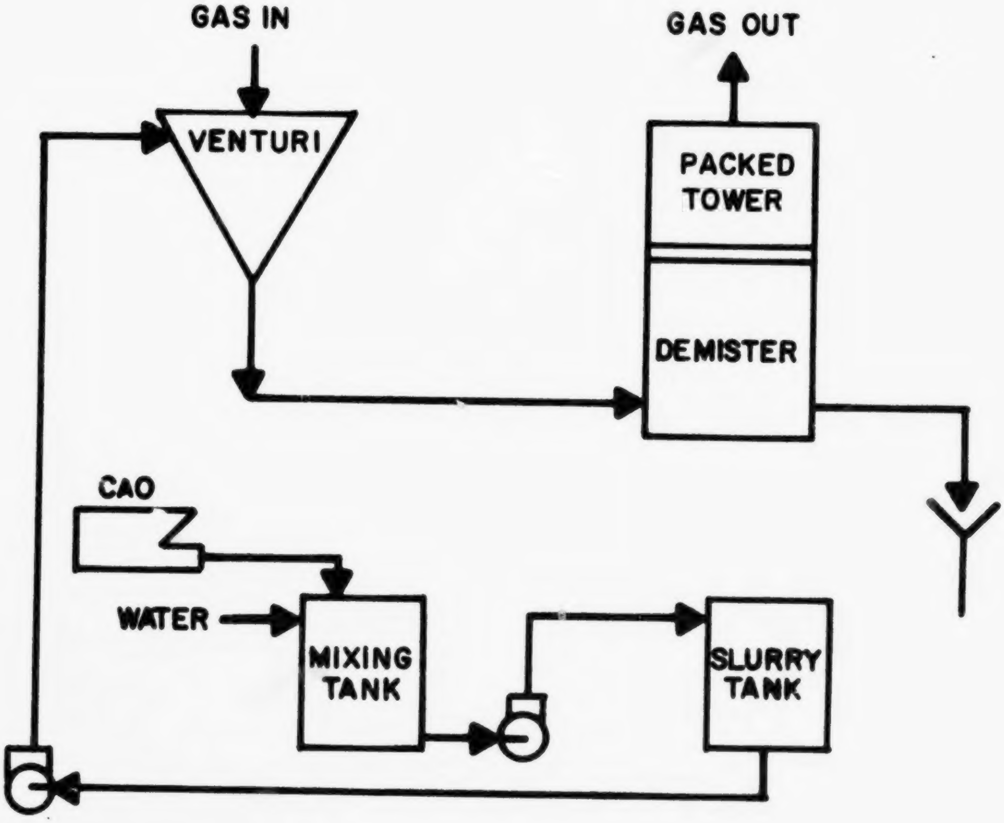

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CAO
WATER
MIXING TANK
SLURRY TANK

FIG. IV-7 CALCIUM OXIDE SLURRY TO FDS - TASK IV

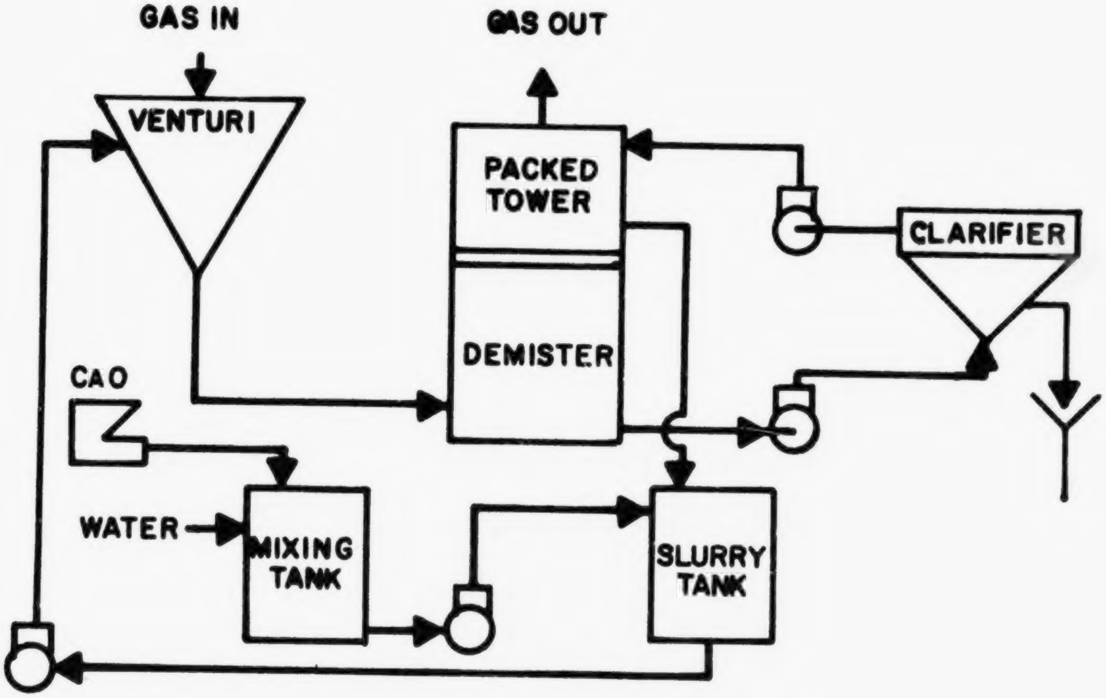

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CLARIFIER
CAO
WATER
MIXING TANK
SLURRY TANK

FIG. IV-8 TWO STAGE CALCIUM OXIDE SCRUBBER WET SLURRY-TASK V

To establish constant operating conditions, the process mode and flows were fixed for several operating days (approximately 8 hours per operating day). Scale buildup within the venturi was severe; the throat disc position (controlling the annular velocity) had to be periodically adjusted to compensate for pressure drop increase. Finally, after 25 hours of operation, the venturi scrubber had to be dismantled and cleaned before the run could be continued.

Results of this test are shown in Table A-6. Pressure drop across the venturi during this test varied from 7 to 22 inches W.G.

Once a stable operation had developed, absorption efficiency ranged from 42 to 58% in the FDS as the pressure drop across the disc increased with a constant L/G of 18 gallons per 1,000 cf. A clarified solution was passed through the tower throughout the run. This had two effects on the tower performance: 1) absorption was limited by restricting the alkali input, and 2) the process was stabilized by eliminating scaling. For a liquid-to-gas ratio of 15 gallons per 1,000 cf, absorption efficiencies of 37 to 63% were measured near the end of the test.

3. Variations in CaO Stoichiometry - Task VIa

Following the above lime slurry run, a CaO dry injection mode was used with the same clarifier solution recirculation to the tower, see Figure IV-9. Venturi scrubber absorption efficiency, 40% removal, was measured at L/G of 18 gallons per 1,000 cf and at a FDS pressure drop of 7.7 inches W.G. Results of these tests are given in Table A-7.

An unusually high tower efficiency was measured during the latter part of these tests; for inlet concentration between 660 to 900 ppm, complete removal of the SO_2 was indicated. Subsequent tests could not reproduce the high efficiency with the same tower feed. If lime slurry passed over from the clarifier, then the high absorption efficiency could be expected.

4. Process Variables Affecting the Venturi Scrubber Performance - Tasks VIb to VIf

The major portion of the test program used a dry injection scheme as shown in Figure IV-10. Again, a clarified

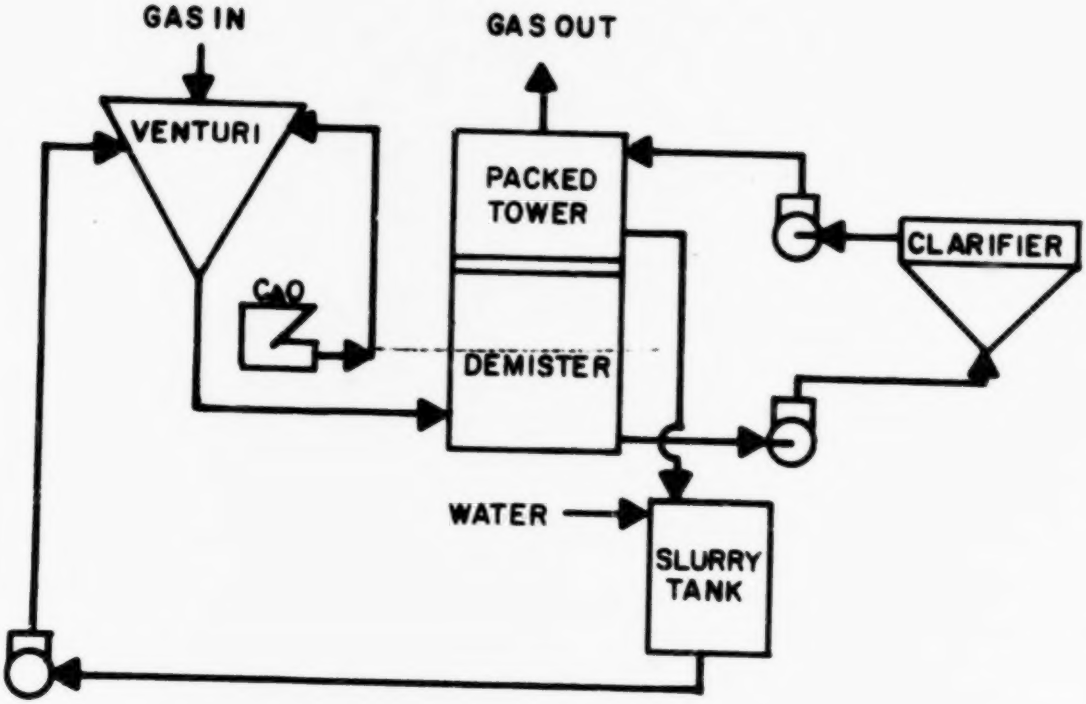

GAS IN
GAS OUT
VENTURI
PACKED TOWER
CLARIFIER
CAO
DEMISTER
WATER
SLURRY TANK

FIG. IV-9 CALCIUM OXIDE DRY INJECTION TO FDS - TASK VIA

solution circulated through the tower while a slurry from the clarifier underflow passed through the FDS. Among the variables studied were stoichiometry, differential pressure, liquid-to-gas ratios, and ionic strengths. The ionic strength was adjusted by the addition of sodium chloride. Concentration of the slurry that passed through the venturi was controlled by the mode illustrated in Figure IV-11. FDS inlet concentration was controlled by proportionating the clarifier underflow and overflow. Results of the tests selected from the main body of data and considered at constant conditions are summarized in Table IV-4.

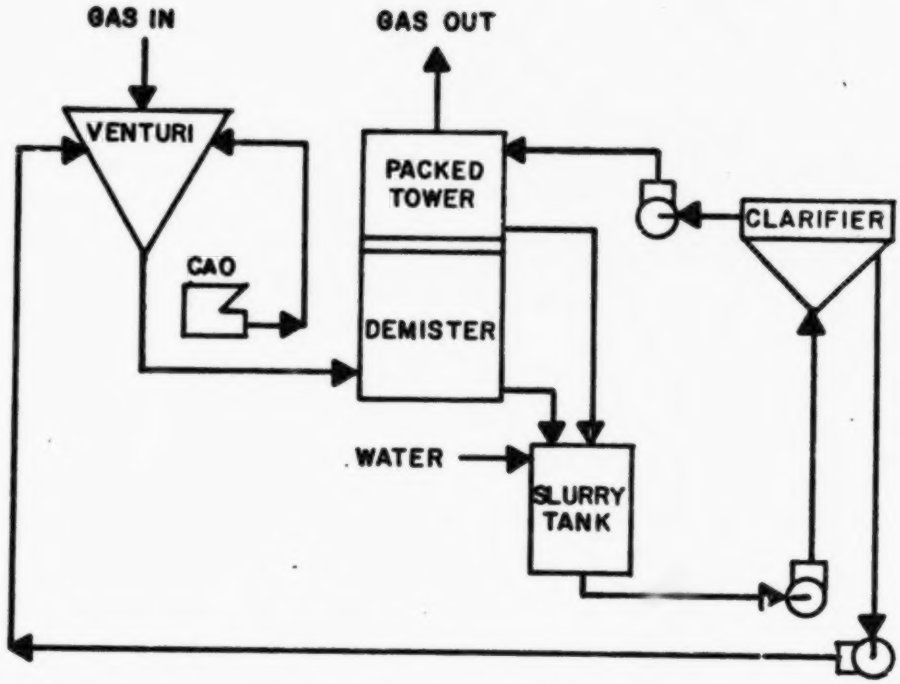

GAS IN
GAS OUT
VENTURI
PACKED TOWER
CLARIFIER
CAO
DEMISTER
WATER
SLURRY TANK

FIG. IV-10 LIME DRY INJECTION TO FDS WITH CLARIFIER RECYCLE- TASKS VI B,C,D&F

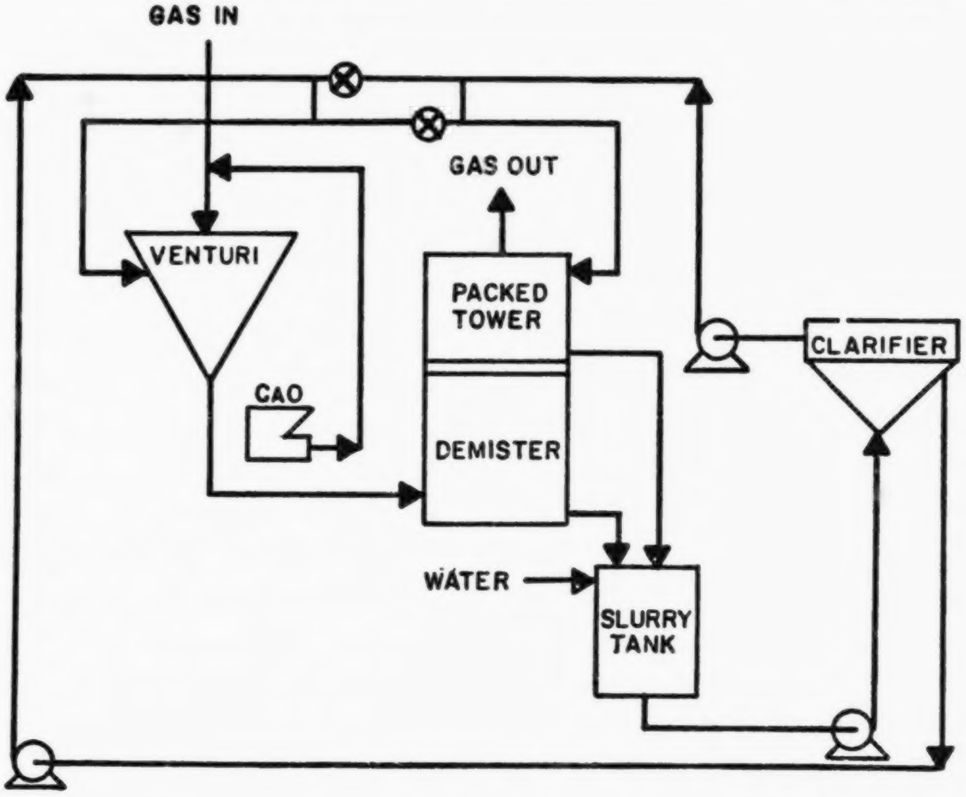

GAS IN
GAS OUT
VENTURI
PACKED TOWER
CLARIFIER
CAO
DEMISTER
WATER
SLURRY TANK

FIG. IV-11 CALCIUM OXIDE DRY INJECTION TO FDS
WITH VARIABLE SLURRY CONCENTRATION TASK VI E

TABLE IV-4

TEST RESULTS USED IN FDS CALCIUM OXIDE CORRELATION

Test No.	Date	Time	SO_2 Collection Efficiency (%)	CaO/SO2 Ratio (Mole/Mole)	Pressure Drop (I.W.C.)	Liquid and Gas Ratio (gal /MCF)	Ionic Strength	Slurry Concentration (%)	mass-transfer coefficient lb.-mole. / (hr)(sq.ft)(atm)
1	2/24	12:15	48.9	1.20	6.0	18.0	0.1	17.5 *	27.5
2	"	13:15	40.4	1.24	6.0	18.0	0.1	17.5 *	21.2
3	"	15:00	34.5	0.97	6.1	10.0	0.1	17.5 *	23.3
4	"	16:00	33.3	.85	5.9	10.0	0.1	17.5 *	22.2
5	"	18:00	46.9	1.05	6.8	23.0	0.1	17.5 *	24.9
6	"	18:30	54.1	1.11	6.8	23.0	0.1	17.5 *	30.6
7	2/25	10:15	42.6	1.10	5.8	18.0	0.1	17.5 *	22.6
8	"	11:25	43.1	1.02	5.8	18.0	0.1	17.5 *	22.9
9	"	12:45	45.5	1.22	5.8	18.0	0.1	17.5 *	24.7
10	"	14:05	45.9	1.04	5.6	18.0	0.1	17.5 *	24.8
11	"	15:35	45.1	1.12	5.6	18.0	0.1	17.5 *	24.2
12	"	16:45	41.8	0.97	6.0	18.0	0.1	17.5 *	22.2
13	4/16	12:30	57.6	1.17	6.8	10.0	1.0	17.5 *	27.9
14	"	13:30	53.8	1.06	7.2	10.0	1.0	17.5 *	43.4
15	"	15:15	54.8	1.09	8.2	10.0	2.0	17.5 *	45.5
16	"	16:15	56.1	1.03	8.6	10.0	2.0	17.5 *	47.5
17	"	19:00	54.3	.9	6.0	10.0	4.0	17.5 *	43
18	"	22:00	53.8	1.04	6.0	10.0	4.0	17.5 *	42.4
19	3/31	12:00	35.7	1.25	6.5	10.0	.1	17.5 *	24.5
20	4/5	15:45	37.5	1.06	12.0	10.3	0.1	17.5 *	28.1
21	4/4	11:30	57.7	1.16	18.2	10.0	0.1	17.5 *	56.2
22	"	19:30	54.3	1.09	19.8	20.0	0.1	17.5 *	43.2
23	4/15	11:45	52.4	0.97	12.7	20.0	0.1	17.5 *	35.8
24	4/1	10:00	41.7	1.0	6.3	10.0	0.1	17.5 *	29.8
25	4/4	10:30	55.2	1.08	18.4	10.0	0.1	17.5 *	52.6
26	"	15:45	56.5	1.03	18.7	20.1	0.1	17.5 *	45
27	4/15	10:15	49.1	1.02	12.5	20.1	0.1	17.5 *	32.4
28	4/1	17:00	41.2	1.01	12.0	10.0	0.1	17.5 *	32.2
29	2/24	17:30	46.8	1.1	6.4	23.0	0.1	17.5 *	24.4
30	5/5	15:00	48.6	1.17	6.4	10.0	0.1	17.5 *	36.8
31	"	16:00	46.7	1.42	7.4	10.0	0.1	3.8	35.5
32	5/6	10:45	50.0	1.03	6.5	10.0	0.1	3.6	38.4
33	"	12:20	48.7	1.11	7.8	10.0	0.1	1.1	38
34	"	13:20	51.1	1.23	7.8	10.0	0.1	1.4	40.7
35	"	15:00	47.8	0.92	7.3	10.0	0.1	1.8	36.6
36	"	10:00	47.4	0.83	7.5	10.0	0.1	5.8	36.3
37	"	17:00	48.2	1.07	7.8	10.0	0.1	6.0	37.4
38	5/13	17:30	38.5	1.0	5.8	10.0	.1	6.7	26.6
39	"	16:30	38.2	0.98	6.1	10.0	.1	16.0	26.5
40	"	15:30	41.3	0.98	6.1	10.0	.1	19.0	29.3
								19.5	

* These slurry concentrations were estimated from the results of Tests 38 through 40.

The underlying purpose of these experiments was the characterization of the flooded disc performance over a wide range of operating conditions. Tower performance, a secondary consideration, was determined for only two slurry conditions and two liquid-to-gas ratios.

5. FDS Results - Task VI

To describe the FDS performance in terms of the key process variables, a regression analysis of 37 sets of selected data was carried out with a computerized multiple linear regression program. Many process models were screened for each of the parameters considered significant. The empirically-derived linear correlation accepted as the best representation of the data for scrubbing with CaO slurry while collecting dry CaO is:

$$\begin{aligned} Y = {} & 29.51 + 5.128\,(R) + 0.983\,(\Delta p) \\ & + 0.701\,L/G + 15.72\,(I) - 2.845\,(I)^2 \\ & - 0.645\,(SL). \end{aligned} \qquad \text{(IV-5)}$$

where Y = SO_2 absorption efficiency of FDS, %,
R = stoichiometric ratio, moles of CaO/mole of SO_2 in inlet,
Δp = pressure drop across throat, inches of H_2O,
L/G = liquid-to-gas ratio, gals./mcf.,
I = ionic strength of NaCl, molarity
SL = slurry concentration, % by weight.

Statistical limitations for this expression are given in Table IV-5.

TABLE IV-5

STATISTICAL PARAMETERS FOR THE CALCIUM OXIDE CORRELATION

Number of Data Sets = 37, See Table IV-4

Correlation Coefficient = 0.862

Standard Error for the Estimate = 3.66%

Significance of Regression (F) = 14.4

Stoichiometric Range, R = 0.83 - 1.42

Pressure Drop, Δp = 5.8 - 19.8 inches of water

Liquid-to-Gas Ratio, L/G = 10 - 23 gals./MCf

Ionic Strength, I = 0.1 - 4

Slurry Concentration, SL = 1.1 - 17.5%

Raw data for the tests used in the correlation are given in Tables A-8 through A-13.

Absorption efficiencies for the lime and sodium carbonate correlations, equations (IV-1) and (IV-5), are compared in Figure IV-12. For low pressure drop, the sodium carbonate and lime showed approximately the same absorption, hence probably the same mass-transfer mechanism. At higher pressure differentials, the lime slurry had a significantly higher absorption resistance.

During all of the lime tests, including some experiments not mentioned thus far, scale buildup in the venturi throat and the scrubber walls caused geometric changes in the absorber. The deposited solids in the throat, as well as the walls, was usually uniform. Normally, a coating of about 1/8 to 1/4 inch occurred very rapidly within the first few hours. The disc had to be adjusted to compensate for the pressure increase resulting from the narrowing down of the throat annulus. Predicting the gas velocity in the throat from the disc position was impossible. A semi-empirical formula describing the venturi pressure drop in a flooded disc scrubber has the general form:*

$$\Delta p = A\ V_t^2\ (L/G + B) \qquad \text{(IV-6)}$$

where A and B = constants,

V_t = throat velocity, ft/sec.,

L/G = liquid-to-gas ratio, gal/mcf.,

Δp = pressure drop across FDS, inches H_2O.

The pressure drop and velocity data for the sodium carbonate tests were used in determining constants A and B.

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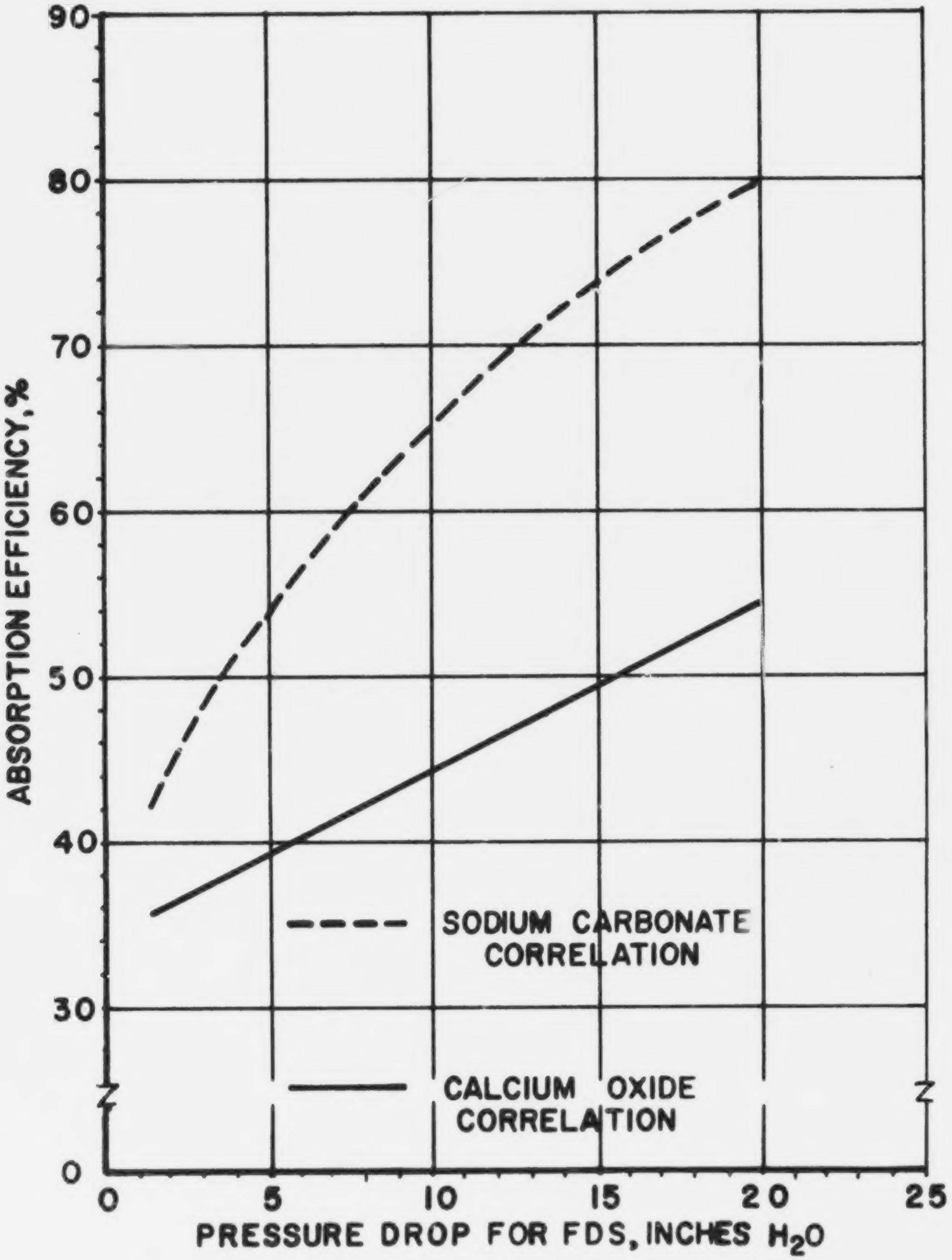
* Robinson, M., "Gas Absorption Mechanisms and Devices, With Special Reference to Flooded-Disc Scrubbers", Project Report PRJ67-9, June 1, 1967.

FIGURE IV-12

ABSORPTION EFFICIENCY FOR CALCIUM OXIDE AND SODIUM CARBONATE AS A FUNCTION OF THE PRESSURE DROP ACROSS THE FDS



The resulting equation is:

$$\Delta p = 3.39 \times 10^{-6} V_t^{2} (L/G + 105) \qquad \text{(IV-7)}$$

Mass-transfer coefficients for the lime absorption were computed from this pressure correlation and equation (II-7). The calculated coefficients for each run are shown in Table IV-4.

6. Power Requirements For Tests VIa Through VId

The estimated horsepower requirements for the lime tests described in sections IV B-2, 3, 4, and 5 are shown in Table A-13. The major portion of the power needed for the SO_2 absorption is in the gas phase; approximately 70 to 85% of the horsepower is consumed by the fan at venturi pressure drop of 6 to 12 inches of H_2O, respectively. For a venturi pressure drop of 6 inches, between 3.2 and 3.8 horsepower are needed per megawatt output on the generator, while at 12 inches Δp on the venturi, 5.3 horsepower per megawatt is used.

7. Effect of Mode Change - Task VII

Three mode changes were made in the process which deviated from the flow patterns used in the developing of equation (IV-5):

1. Lime Feed via Slurry without dry injection
2. Slurry Feed to Tower and FDS
3. Clarified Solution to Tower and FDS.

With these mode changes, the absorption effects for either slurry liquors and/or clarified solution could be isolated. Operating conditions and absorption efficiency measurements for each of the tests are discussed below in section B-8, 9, and 10. Detailed data on each test run are given in Tables A-14 through A-16.

8. Lime Feed via Slurry Without Dry Injection - Task VIIa

In this mode, a 1.0% lime slurry was passed from the mixing tank to the hold tank where it was mixed with the venturi discharge. A slurry blowdown from the clarifier was circulated through the venturi while the clarified overflow was sent through the tower. Tower discharge bypassed the hold tank and entered directly into the clarifier, see Figure IV-13.

FDS absorption efficiency for this test was higher than achieved for the same operating conditions predicted with equation (IV-5). Between 48 to 53% of the inlet SO_2 was removed with a pressure differential across the disc of 6.5 to 8.5 inches W.G. and a L/G of 10 gallons per 1,000 cf. Slurry temperature to the disc for this run measured 100 to 102°F or about 20 to 40°F lower than normal. This lower temperature may have influenced the absorption by increasing lime solubility and reducing the SO_2 vapor pressure.

Results of the test are compared to equation (IV-5) in Figure IV-14. Approximately 10 to 15% more absorption took place for this test than predicted by the correlations. Detailed operating conditions for the test are presented in Table A-14.

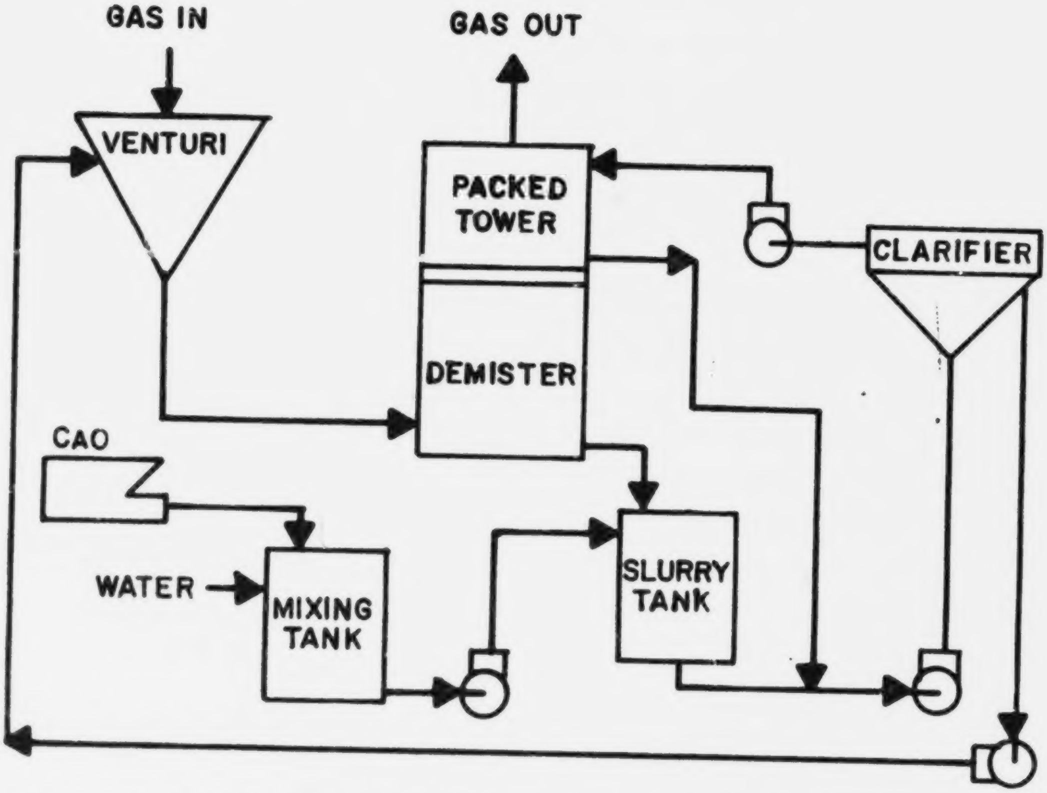
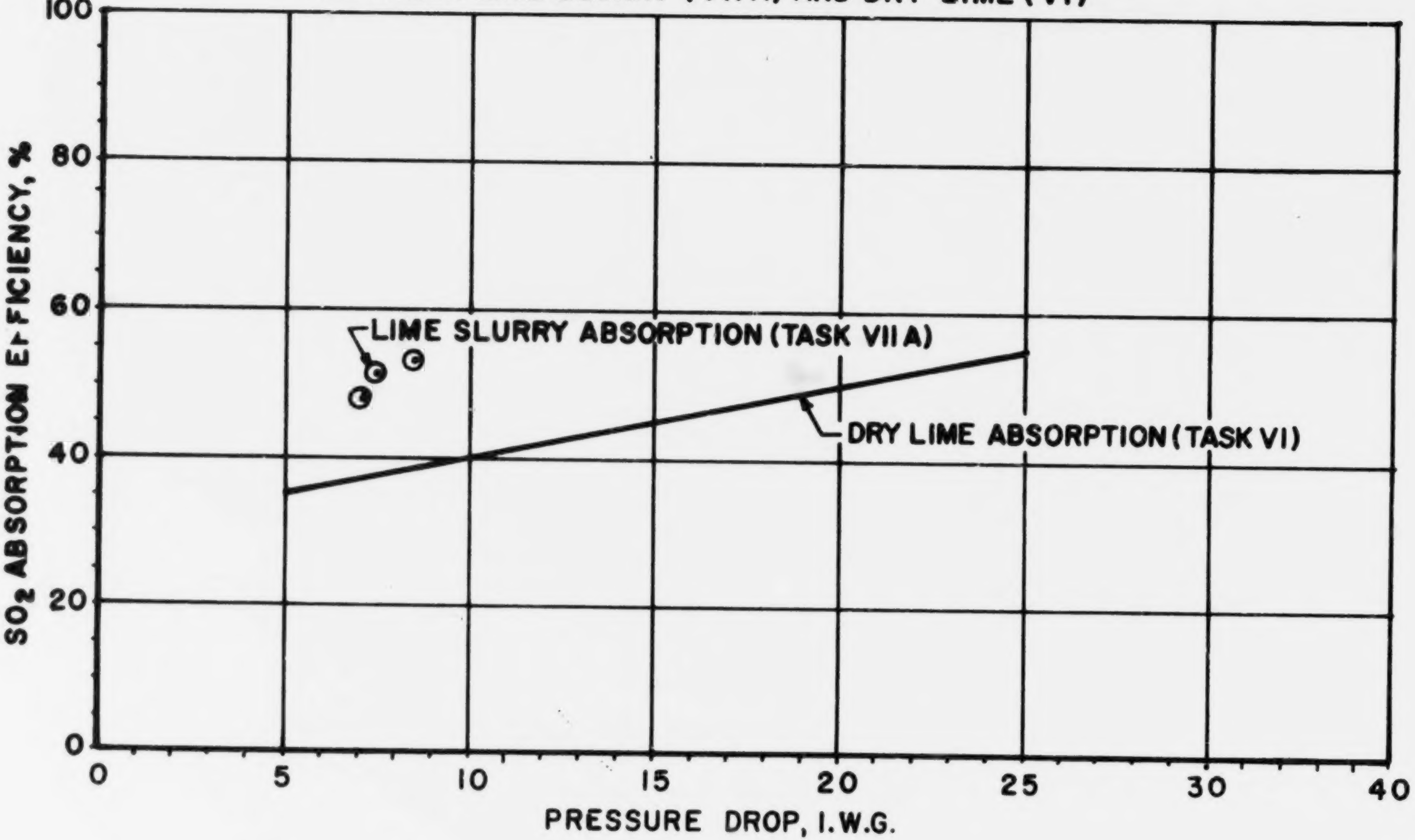

GAS IN
GAS OUT
VENTURI
PACKED TOWER
CLARIFIER
DEMISTER
CAO
WATER
MIXING TANK
SLURRY TANK

FIG. IV-13 WET SLURRY TO FDS WITH CLARIFIER RECYCLE-TASK VIIA

FIGURE IV-14
COMPARISON OF SO_2 ABSORPTION VS, PRESSURE DROP BETWEEN LIME SLURRY (VII A) AND DRY LIME (VI)



9. Slurry Feed to the Venturi and Tower - Task VIIb

For this test, slurry containing calcium oxide, reaction products, and fly ash circulated through both the tower and the FDS, as illustrated in Figure IV-15. The solids concentration to the venturi was held at 15 to 19% while the tower slurry varied from 3.4 to 4.4%. Process conditions remained constant throughout the run at L/G = 10 gallons per 1,000 cf for the FDS and 20 gallons per 1,000 cf for the tower.

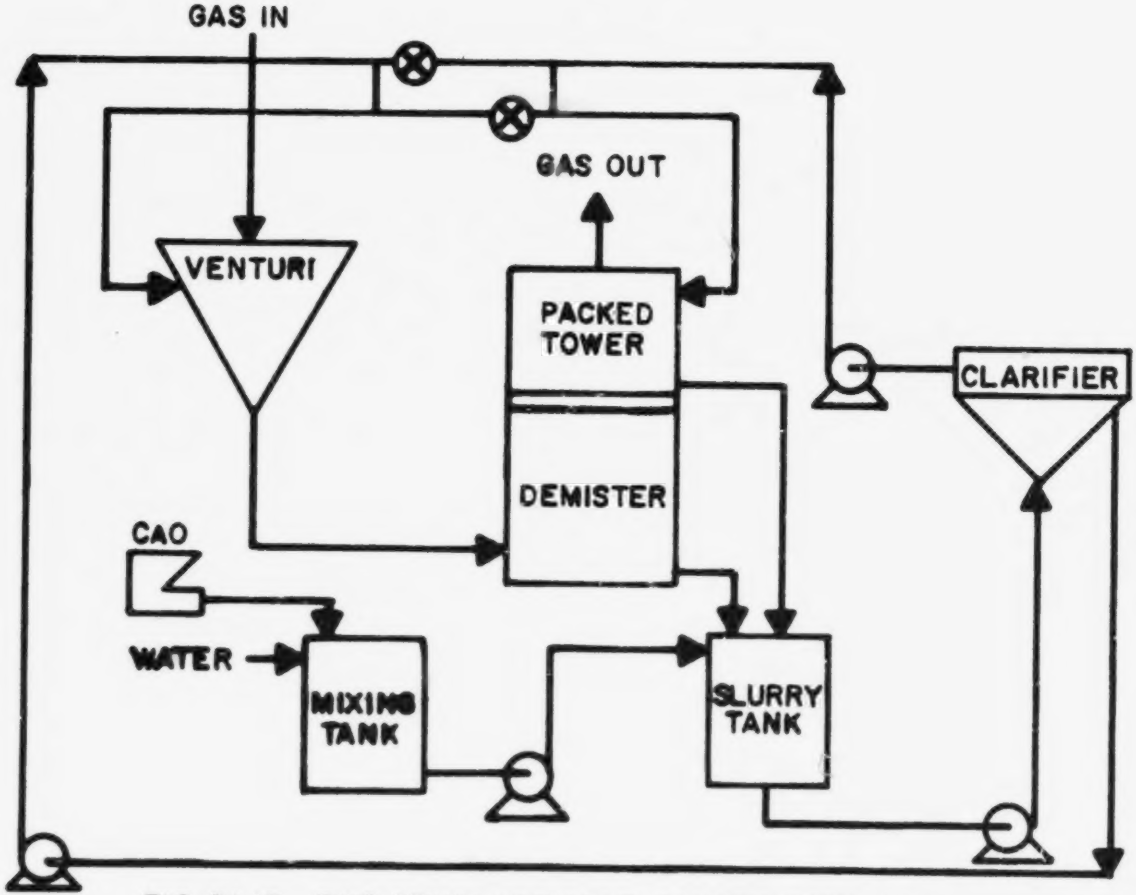

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CLARIFIER
CAO
WATER
MIXING TANK
SLURRY TANK

FIG. IV-15 TWO STAGE VARIABLE SLURRY-TASK VIIB

Results of these tests are given in Table A-15. The efficiency for the FDS (approximately 39%) was about 10% less than determined with the lower slurry concentration predicted by equation (IV-5). This data was added to the results used in equation (IV-5) and a slightly improved regression was derived, i.e.

$$Y = 30.7 + 4.57\ (R) + 0.952\ (\Delta p) + 0.647\ (L/G) + 15.16\ (I) - 2.751\ (I)^2 - 0.598\ (SL) \qquad \text{(IV-8)}$$

where Y = SO_2 absorption efficiency of FDS, %,

R = stoichiometric ratio, moles of CaO/mole of SO_2 in inlet,

Δp = pressure drop across disc, inches of H_2O,

L/G = liquid-to-gas ratio, gals/mcf.,

I = ionic strength of NaCl, molarity

SL = slurry concentration, % by weight.

Statistical parameters for this correlation are given in Table IV-6.

10. Clarified Solution to Tower and FDS - Task VIIc

The objective of this test was the determination of the FDS absorption efficiency without solid suspension. A clarified solution was pumped to both the FDS and the tower in the mode given in Figure IV-16. SO_2 removal, as expected, dropped off considerably. Between 13 to 15% of the SO_2 was absorbed in the FDS for inlet gas concentration at 1,560 to 1,650 ppm. pH of the lime solution varied between 11.1 to 5.2

TABLE IV-6

STATISTICAL PARAMETERS FOR THE CALCIUM OXIDE CORRELATION

Number of Data Points = 40, See Table IV-4

Correlation Coefficient = 0.873

Standard Error of the Estimate = 3.57%

Significance of Regression (F) = 17.6

Stoichiometric Range, R = 0.83 - 1.42

Pressure Drop Range, Δp = 5.6 - 19.8 inches of water

Liquid-to-Gas Ratio Range, L/G = 10 - 23 gals/MCF

Ionic Strength Range, I = 0.1 - 4

Slurry Concentration Range, SL = 1.1 - 19.5%

as it passed through the FDS; hence, a major portion of the alkali was consumed. The dissolved lime contributes only a small fraction of the required absorbate. Results of this test are summarized in Table A-16.

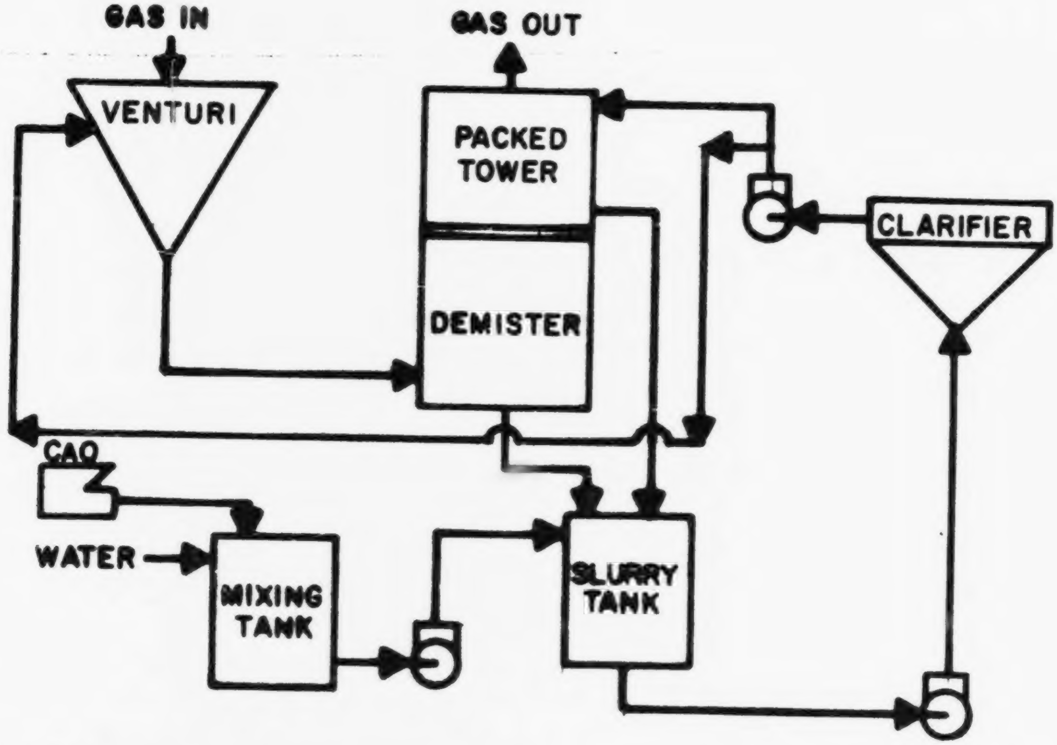

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CLARIFIER
CAO
WATER
MIXING TANK
SLURRY TANK

FIG. IV-16 WET SLURRY WITH CLARIFIER OVERFLOW TO FDS-TASK VIIC

11. Tower Absorption

Operating parameters for the tower were deliberately fixed at one level for most of the test program to minimize

FDS process variations. Yet, efficiency for the tower ranged from 24 to 96%. Uncontrollable process conditions such as inlet SO_2 concentration, gas dew point temperature, and inlet liquid alkalinity, varied considerably. An attempt to correlate the tower absorption efficiency data with the variations that did occur did not yield any meaningful mathematical expression. A weak absorption efficiency relationship with inlet gas concentration, liquid phase temperature, and solution pH was evident.

For the majority of tests, a clarified solution passed through the tower. Solution pH entering the absorber at pH 10 to 11.2 changed considerably as it absorbed the SO_2. Exit pH varied between 2.0 and 9.4. Estimated stoichiometry for the input SO_2 and $Ca(OH)_2$ indicated a limiting alkalinity when a saturated clarified solution was fed; however, for most tests, minor solution turbidity was observed at the clarifier overflow. The alkaline nature of the solids creating the turbidity could affect the stoichiometry significantly. Absorption efficiency measurements for a major portion of the calcium oxide tests are presented in Table IV-7.

Solution ionic strength was varied by the addition of sodium chloride for three levels. The object of these tests was the simulation of steady-state conditions where chloride ion and sodium ion would build up in a closed-loop process. The three levels of concentration selected, 1.0, 2.0, and 4.0 molality, indicated a maximum absorption at I = 2. At an ionic strength of 1, the absorption ranged from 46 to 65% while at I = 2, the absorption was between 91 and 94%. Increasing the sodium chloride concentration to I = 4 adversely affected the absorption; and between 79 and 91% removal was observed. A

TABLE IV-7

SELECTED RUN DATA FOR THE PACKED TOWER

Test No.	Date	Time	SO_2 Collection Efficiency (%)	SO_2 Inlet Loading (ppm)	Liquid & Gas Ratio (gals/cfm)	Slurry Concentration* (%)	Slurry Temp. of Clarifier (°F.)	Ionic Strength	pH of Tower Feed	pH of Tower Underflow	Gas Temp. to Tower (°F.)	Gas Temp. From Tower (°F.)	Mass-Transfer Coefficient lb-mols (hr.)(sq.ft.)(atm)
1	2/24	12:15	72.6	690.	15.	0.1	138.	0.1	10.7	5.2	150.	140.	0.32
2	2/24	13:15	45.3	840.	15.	0.1	145.	0.1	10.4	5.2	152.	148.	0.15
3	2/24	15:00	37.0	1080.	15.	0.1	148.	0.1	10.4	5.0	152.	150.	0.11
4	2/24	16:00	33.2	1200.	15.	0.1	140.	0.1	10.6	5.1	152.	148.	0.10
5	2/24	17:00	35.3	990.	15.	0.1	120.	0.1	10.6	5.7	140.	135.	0.11
6	2/24	18:00	23.5	1020.	15.	0.1	135.	0.1	10.6	5.8	150.	148.	0.07
7	2/24	18:30	32.3	945.	15.	0.1	136.	0.1	10.6	5.7	150.	140.	0.10
8	3/30	14:00	89.8	960.	15.	0.1	95.	0.1	11.8	4.2	115.	100.	0.58
9	3/30	16:30	55.6	810.	15.	0.1	105.	0.1	11.2	4.8	118.	108.	0.20
10	3/31	10:00	51.4	1110.	15.	0.1	102.	0.1	11.1	4.7	118.	102.	0.18
11	3/31	12:00	52.8	1080.	15.	0.1	108.	0.1	11.1	4.8	120.	110.	0.19
12	4/1	10:00	49.1	1120.	15.	0.1	100.	0.1	11.1	4.0	118.	108.	0.17
13	4/1	12:00	48.7	1170.	15.	0.1	110.	0.1	11.0	4.1	122.	115.	0.17
14	4/1	14:00	44.4	1080.	15.	0.1	116.	0.1	11.2	4.1	122.	118.	0.15
15	4/1	17:00	36.7	1200.	15.	0.1	109.	0.1	11.3	5	120.	105.	0.12
16	4/5	15:[illegible]5	51.7	1200.	15.	0.1	104.	0.1	11.4	4.5	112.	105.	0.17
17	4/7	11:30	55.6	810.	15.	0.1	110.	0.1	9.9	3.6	115.	105.	0.2
18	4/7	13:30	64.8	415.	15.	0.1	112.	0.1	11.3	3.9	118.	108.	0.25
19	4/7	16:00	68.0	363.	15.	0.1	112.	0.1	11.2	4.5	118.	110.	0.28
20	4/7	20:00	57.8	450.	15.	0.1	108.	0.1	11.2	3.8	115.	100.	0.18
21	4/14	10:30	90.3	780.	15.	0.1	111.	0.1	11.0	4.0	115.	107.	0.57
22	4/14	11:30	94.0	705.	15.	0.1	111.	0.1	11.0	4.5	117.	109.	0.69
23	4/14	12:40	95.6	720.	15.	0.1	112.	0.1	11.0	4.7	118.	112.	0.77
24	4/14	15:45	88.0	600.	15.	0.1	103.	0.1	10.9	4.2	112.	108.	0.52
25	4/14	18:00	51.6	620.	15.	0.1	108.	0.1	10.5	4.6	120.	110.	0.18
26	4/14	19:00	52.4	630.	15.	0.1	108.	0.1	10.5	4.5	118.	112.	0.18
27	4/14	22:20	30.6	735.	15.	0.1	108.	0.1	10.4	4.0	118.	112.	0.09

* Some minor solution turbidity was evident throughout the test program.

TABLE IV-7 cont'd

SELECTED RUN DATA FOR THE PACKED TOWER

Test No.	Date	Time	SO_2 Collection Efficiency (%)	SO_2 Inlet Loading (ppm)	Liquid & Gas Ratio (gals/cfm)	Slurry Concentration (%)	Slurry Temp. of Clarifier (°F.)	Ionic Strength	pH of Tower Feed	pH of Tower Underflow	Gas Temp. to Tower (°F.)	Gas Temp. From Tower (°F.)	Mass-Transfer Coefficient lb-mols / (hr.)(sq.ft.)(atm
28	4/15	10:15	30.8	780.	15.	0.1	114.	0.1	10.8	4.2	115.	113.	0.09
29	4/15	11:45	29.6	750.	15.	0.1	109.	0.1	11.1	4.2	115.	111.	0.09
30	4/15	12:45	41.6	945.	15.	0.1	109.	0.1	11.1	4.2	115.	111.	0.13
31	4/15	15:00	52.4	870.	15.	0.1	107.	0.1	11.3	4.4	115.	111.	0.19
32	4/15	16:30	54.9	825.	15.	0.1	107.	0.1	11.3	4.5	115.	110.	0.21
33	4/15	17:30	57.2	780.	15.	0.1	107.	0.1	11.3	4.4	118.	110.	0.21
34	4/15	20:30	52.3	810.	15.	0.1	110.	0.1	11.0	4.8	118.	112.	0.18
35	4/15	22:00	51.1	810.	15.	0.1	108.	0.1	11.0	4.8	118.	112.	0.18
36	2/25	10:15	53.1	1110.	15.	0.1	118.	0.1	10.0	5.8	138.	125.	0.19
37	2/25	11:20	40.6	1110.	15.	0.1	132.	0.1	10.8	6.2	144.	134.	0.13
38	2/25	12:45	57.7	900.	15.	0.1	140.	0.1	11.3	5.9	142.	142.	0.21
39	2/25	14:05	57.7	900.	15.	0.1	138.	0.1	11.3	6.0	138.	137.	0.21
40	2/25	15:35	64.0	840.	15.	0.1	130.	0.1	11.4	5.9	138.	135.	0.25
41	2/25	16:45	54.2	960.	15.	0.1	115.	0.1	11.3	5.8	132.	128.	0.19
42	5/5	9:30	60.9	1065.	15.	0.1	96.	0.1	11.0	3.6	122.	106.	0.24
43	5/5	10:30	79.0	1050.	15.	0.1	106.	0.1	11.0	3.8	116.	105.	0.39
44	5/5	11:15	73.9	1020.	15.	0.1	102.	0.1	11.0	3.8	116.	108.	0.34
45	5/5	13:30	63.6	1320.	15.	0.1	104.	0.1	11.1	2.4	118.	112.	0.25
46	5/5	15:00	73.7	1110.	15.	0.1	110.	0.1	11.2	2.3	120.	115.	0.34
47	5/5	16:00	86.5	960.	15.	0.1	112.	0.1	11.3	2.2	120.	116.	0.50
48	5/6	10:45	82.6	1010.	15.	0.1	102.	0.1	11.4	2.0	118.	110.	0.44
49	5/6	12:20	76.0	1015.	15.	0.1	102.	0.1	11.0	2.0	118.	110.	0.36
50	5/6	13:20	76.0	870.	15.	0.1	104.	0.1	11.2	2.2	118.	110.	0.36
51	5/6	15:00	86.0	930.	15.	0.1	106.	0.1	11.3	2.4	120.	112.	0.49
52	5/6	16:00	83.3	900.	15.	0.1	106.	0.1	11.3	2.2	118.	112.	0.45
53	5/6	17:00	85.1	870.	15.	0.1	108.	0.1	11.2	2.2	118.	114.	0.48

TABLE IV-7 cont'd

SELECTED RUN DATA FOR THE PACKED TOWER

Test No.	Date	Time	SO_2 Collection Efficiency (%)	SO_2 Inlet Loading (ppm)	Liquid & Gas Ratio (gals/cfm)	Slurry Concentration (%)	Slurry Temp. of Clarifier (°F.)	Ionic Strength	pH of Tower Feed	pH of Tower Underflow	Gas Temp. to Tower (°F.)	Gas Temp. From Tower (°F.)	Mass-Transfer Coefficient lb-mols (hr.)(sq.ft.)(atm)
54	4/16	11:30	46.1	600.	15.	0.1	110.	1.0	11.2	4.4	115.	113.	0.16
55	4/16	12:30	63.7	750.	15.	0.1	109.	1.0	11.2	5.8	115.	110.	0.26
56	4/16	13:30	65.3	900.	15.	0.1	106.	1.0	11.2	5.8	115.	115.	0.27
57	4/16	15:15	94.3	840.	15.	0.1	106.	2.0	11.2	7.0	114.	110.	0.72
58	4/16	16:15	91.5	870.	15.	0.1	108.	2.0	11.2	6.6	114.	110.	0.63
59	4/16	19:00	91.0	900.	15.	0.1	108.	4.0	10.8	9.4	118.	118.	0.59
60	4/16	20:00	79.0	570.	15.	0.1	108.	4.0	10.9	6.5	118.	118.	0.39
61	4/16	22:00	79.1	630.	15.	0.1	110.	4.0	11.0	6.2	118.	118.	0.39
62	5/13	14:00	79.1	1420.	20.	4.4	98.	0.1	11.1	3.8	112.	100.	0.4
63	5/13	15:30	84.5	1125.	20.	3.4	102.	0.1	11.1	2.5	113.	103.	0.48
64	5/13	16:30	77.1	1260.	20.	3.5	102.	0.1	11.1	2.4	115.	106.	0.38
65	5/13	17:30	85.9	1110.	20.	3.4	104.	0.1	11.1	2.2	115.	108.	0.5

summary of the tower conditions for the ionic strength experiment is shown in Table IV-7, Tests 54 through 61.

For Tests 55 and 56, ionic strength = 1, the amount of SO_2 being absorbed per unit time was compared to the hydroxide solubility predicted with an equilibrium model.[9] The consumed hydroxide calculated from the absorbed SO_2 and solution flow was very close to the theoretical saturation solubility of the liquid. With ionic strength equal to 1.055, the theoretical $[OH]^-$ concentration is 1.133×10^{-2} g-moles/liter which is almost exactly the predicted hydroxide consumption of 1.24×10^{-2} or 1.44×10^{-2} g-moles/liter for tests 55 and 56 respectively. These results are summarized in Table IV-8.

To determine the SO_2 absorption during slurry feed to the tower, several tests were performed with slurry inputs ranging from 3.4 to 4.4% solids. Absorption measured between 77 and 86% with inlet SO_2 concentrations ranging from 1110 to 1420 ppm. Results of these experiments are listed in Tests 62 to 65 of Table IV-7. No outstanding absorption improvement could be seen with the high slurry feed (3.4 to 4.4%) compared to the efficiency measurements with clarified solution.

Tower mass-transfer coefficients ranged from 0.1 to 0.77 lb-moles/(hr)(sq.ft.)(atm) for the tests shown in Table IV-7. High ionic strength solutions and slurry feed gave mass-transfer coefficients between 0.4 and 0.7.

C. OTHER ALKALI MATERIALS - TASK VIII

Thus far the absorption measurements have simulated the Dry Injection-Wet Scrubbing Process with a soft burned calcium oxide. In the next section, other limestone materials

TABLE IV-8

The Comparison Between the Solubility Data from Radian Corporation [1] and Data from the Efficiency Measurement

	SO_2 Removed Efficiency in Tower (%)	Tower SO_2 Inlet loading (PPM)	Tower Feed Temp. (° F)	Tower Feed PH Value	Ionic Strength	Conc. of [OH] in Tower Feed gr.-mols/liter
Radian's Data	-	-	131	11.214	1.055	1.133×10^{-2}
Measured Data						
(2)	63.7	750	109	11.2	1.	1.24×10^{-2} (4)
(3)	65.3	900	106	11.2	1.	1.44×10^{-2} (4)

---

(1) Radian Corporation: A Theoretical Description of the Limestone Injection-Wet Scrubbing Process, Volume 11, B-2, (1970).

(2) Table 4- Test #55

(3) Table 4- Test #56

(4) Hydroxide consumed in the Tower by the absorbed SO_2

were processed for comparison. A high grade dolomitic lime, containing approximately 99% MgO.CaO and a partially sulfated lime/fly ash material having 25.3% CaO were processed with mode conditions similar to the calcium oxide tests.

Specific operating conditions for these tests (section IV C1, 2, and 3) are summarized in Tables A-17 through A-19.

1. Dolomitic Lime - Task VIIIc

Task VIIIc, absorption efficiency measurements with dolomitic lime (CaO.MgO), were performed with the operational scheme illustrated in Figure IV-17. Venturi scrubber absorption efficiencies for this "once through" process ranged from 56 to 69% removal with a FDS Δp between 7.0 and 8.2" of W.G. The dolomitic lime gave an efficiency 27 to 37% higher than measured with the calcined limestone. With calcium oxide, the absorption efficiency for a similar operation (Task III) was 30% for a stoichiometric ratio of 1.0 and a $\Delta p = 10$" W.G. These results are compared in Figure IV-18. Stoichiometric ratio for the dolomitic lime was computed with the CaO plus the MgO.

2. Sulfated Lime/Fly Ash via Dry Injection - Task VIIId

Partially sulfated lime/fly ash material from the Shawnee Power Station, Paducah, Kentucky was processed in the pilot system using again the operating mode illustrated in Figure IV-17. The lime/fly ash material tested contained 25.3% free CaO. Under these conditions, fly ash loading was appreciably higher than usual. Although absorption efficiency was low, there is no evidence that fly ash concentration was directly responsible. Absorption efficiencies of 11.6 to 18.4% were measured

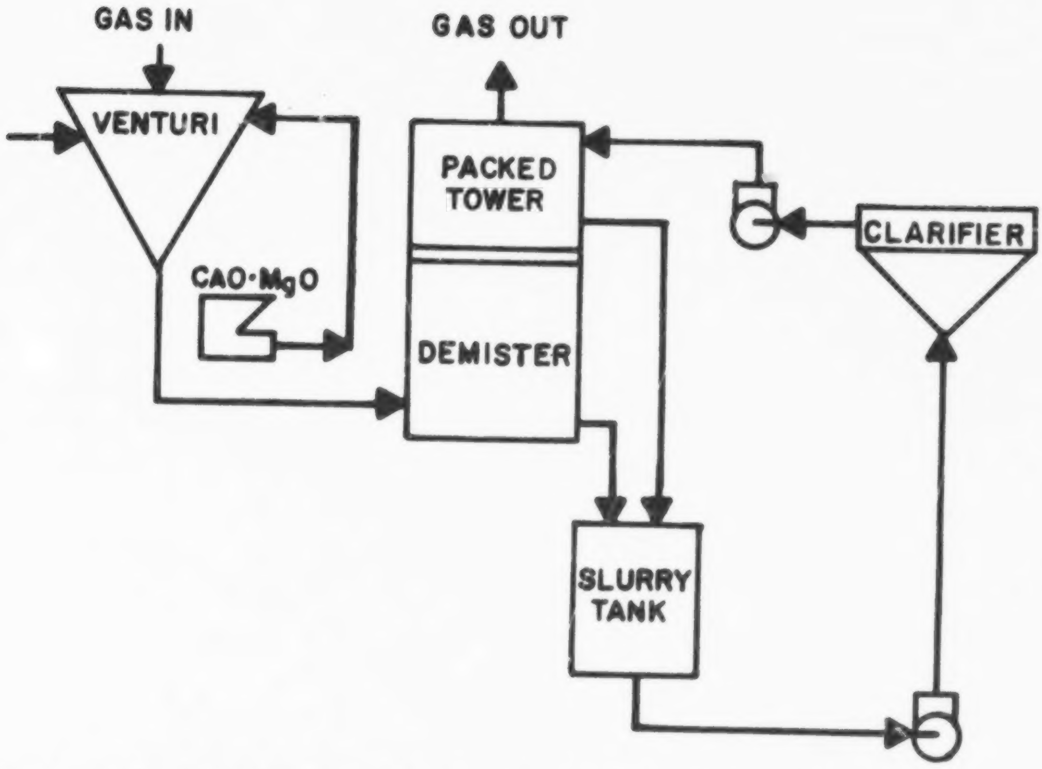

GAS IN
GAS OUT
VENTURI
PACKED TOWER
CLARIFIER
CAO·MgO
DEMISTER
SLURRY TANK

FIG. IV-17 DRY INJECTION OF DOLOMITIC LIME TO FDS-TASK VIIIC&D

for an L/G = 10 gallons per 1,000 cf and a pressure drop ranging from 5.7 to 10.8 inches of water. Stoichiometric ratio, as given in Table A-18, was controlled at 0.93 to 1.02.

Results of this test are graphically compared in Figure IV-18 to calcium oxide and dolomitic lime. As the graph indicates, the sulfated lime performed poorly in comparison to both calcium oxide and dolomitic lime under similar operating conditions. With dry injection, calcium oxide removed 30% of the inlet SO_2 while the sulfated lime/fly ash absorbed approximately 17%.

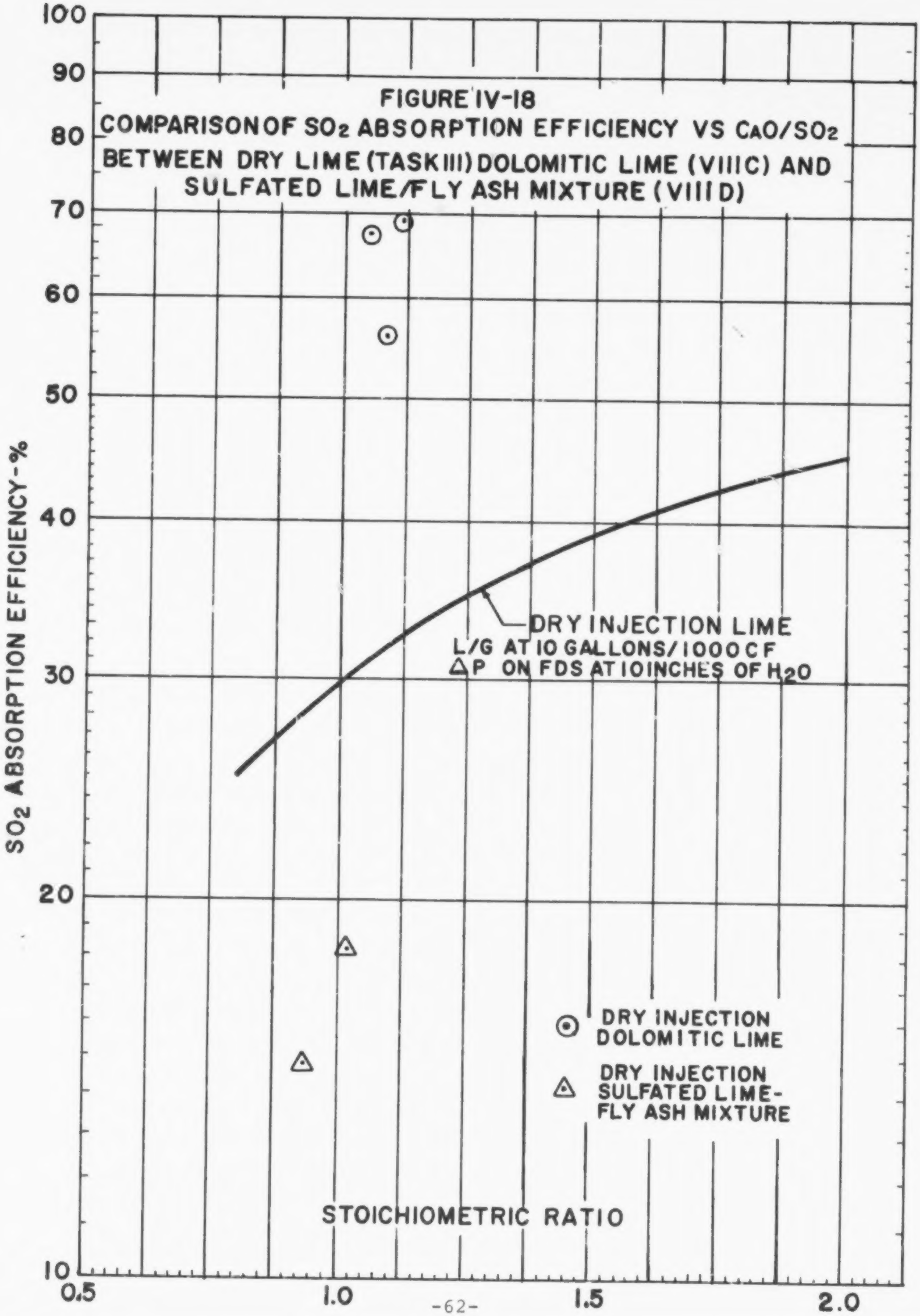


FIGURE IV-18
COMPARISON OF SO_2 ABSORPTION EFFICIENCY VS CaO/SO_2 BETWEEN DRY LIME (TASK III) DOLOMITIC LIME (VIII C) AND SULFATED LIME/FLY ASH MIXTURE (VIII D)

Pressure drop across the disc increased from 6.8 to 10.8 inches W.G. in 4 hours of operation while the throat setting remained in the opened position. This increase in Δp would indicate a significant buildup of solids within the throat area.

Absorption in the packed tower with a clarified solution input was between 37 and 50% removal for L/G = 20 gallons per 1,000 cf. The efficiency for the tower did change during the process run and its final value was 37.5%. This low efficiency is predictable from the pH values (6.8 to 8.6) for the tower slurry tank inlet in Table A-18. Calcium oxide for the same process conditions, Task VIb, demonstrated similar low absorption with a clear liquid feed, i.e. 25 to 35% removal as given in Table A-8.

3. Sulfated Lime/Fly Ash With Wet Slurry - Task VIIIe

The operating mode for this task is shown in Figure IV-19. A 4% slurry of lime/fly ash was passed from the mix tank to the slurry tank feeding both the venturi and the tower. The clarifier was bypassed to allow a slurry input to the tower.

FDS sulfur dioxide absorption was lower than the lime for a comparable mode. These results are compared in Figure IV-20. Here, the SO_2 removal efficiency for the 4% slurry measured approximately 20 to 31% for the FDS while the calcium oxide, at the same process conditions, allowed 47% removal. Residence times in the slurry tank were approximately the same for the calcium oxide and sulfated lime/fly ash, i.e. 40 to 60 minutes.

Tower efficiency for the 4% sulfated lime/fly ash slurry was outstandingly good; 93 to 97% SO_2 removal was established

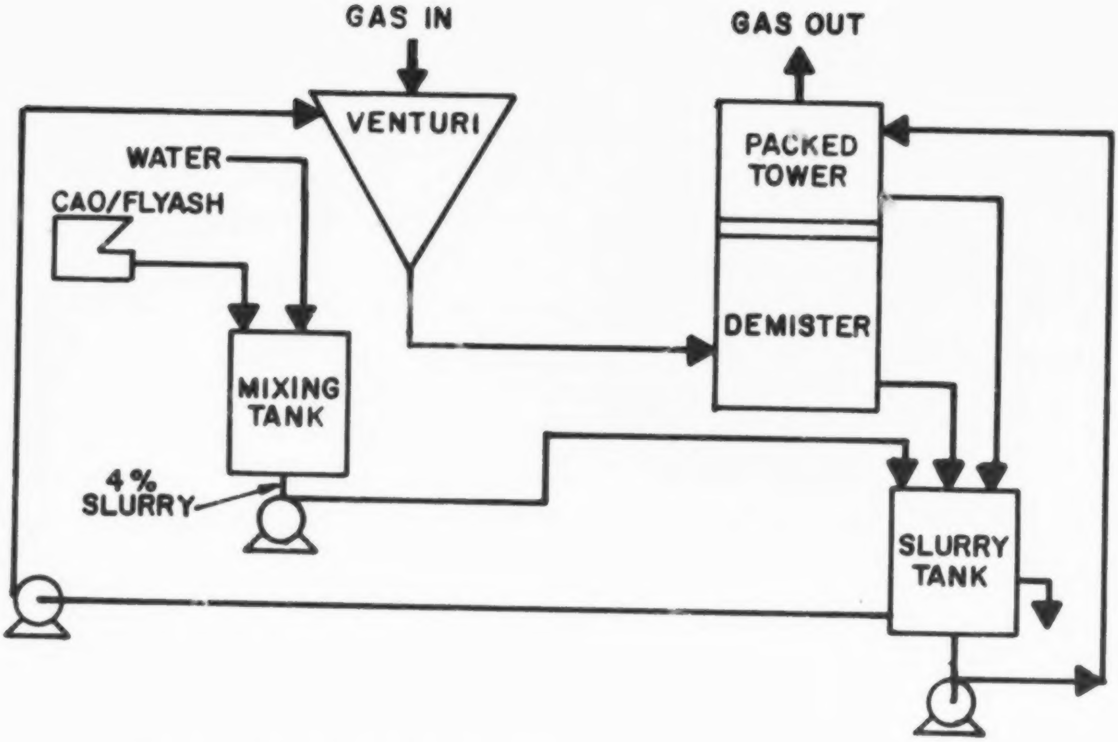

GAS IN
GAS OUT
VENTURI
PACKED TOWER
WATER
CAO/FLYASH
DEMISTER
MIXING TANK
4% SLURRY
SLURRY TANK

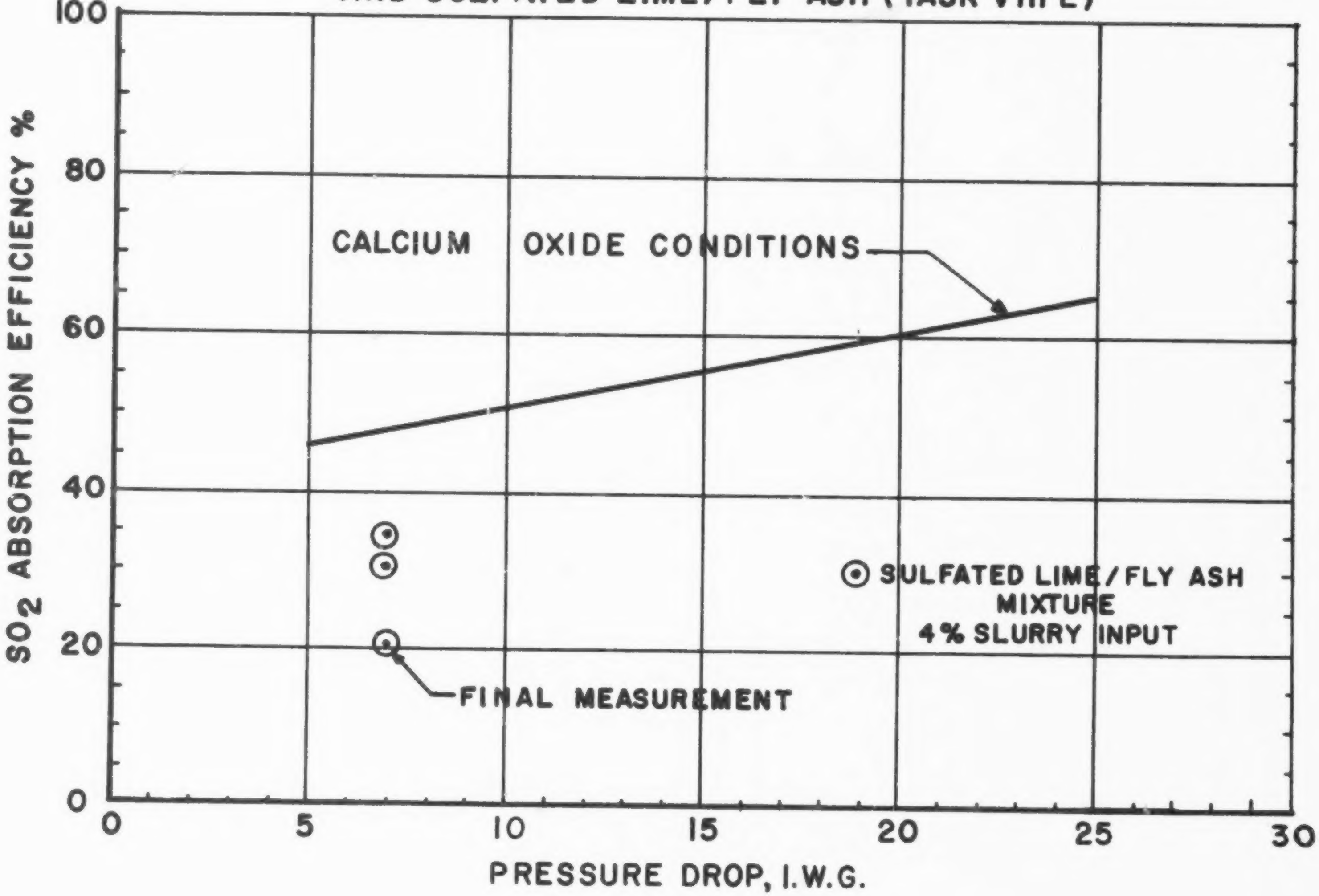
FIG. IV-19 LIME/FLYASH SLURRY TO FDS-TASK VIII E

at a L/G ratio of 20 gallons per 1,000 cf. Calcium oxide at similar conditions, 3.4 to 4.4% solids, gave lower absorption, i.e. between 77 and 87% removal. The sulfated lime/fly ash was well mixed in the slurry tank before entering the tower while the calcium oxide test used clarifier blowdown. The sulfated lime presumably dissolved to a greater extent in the agitated vessel than the calcium oxide did in the large stagnant clarifier.

Detailed operating conditions for the sulfated lime tests are given in Table A-19.

FIG. IV-20

COMPARISON OF SO_2 ABSORPTION EFFICIENCY VS. PRESSURE DROP BETWEEN CALCIUM OXIDE (TASK VI & VII) AND SULFATED LIME/FLY ASH (TASK VIII E)



D. LIMESTONE

The limestone programs were carried out in two separate test series in cooperation with the Tennessee Valley Authority. In the first program, several types of limestone and one hydrated lime were processed in an open-loop system; absorption efficiencies were compared for each alkali type. Following these tests, a second investigation was performed studying the scale accumulation within the tower absorber using a limestone selected from the first test run.

1. Limestone Efficiency Tests - Open-Loop

Eight absorption tests were executed with four carbonate compounds, one hydrated lime, two liquid-to-gas ratios, and two operating modes. The four calcium carbonate materials were provided by TVA; a list of these materials and their chemical analyses are given in Table B-1, Appendix B.

Block diagrams illustrating the two operating modes are shown in Figure IV-21. For the major portion of these experiments, reaction slurry was fed to both the FDS and the packed tower. Alkali slurry flowed countercurrently to the gas as it passed from the tower slurry tank to the venturi slurry tank. A two percent by weight $CaCO_3$ or $Ca(OH)_2$ slurry was pumped from the mixing tank to the tower slurry tank at the stoichiometric rate. An equal slurry flow passed from the tower tank to the venturi slurry tank where it then overflowed to the discharge.*

Hydrated lime, the most efficient alkali of the group, removed 99% of the SO_2. Limestone and chalk gave an efficiency of 96% while cement dust, the least effective material, scrubbed

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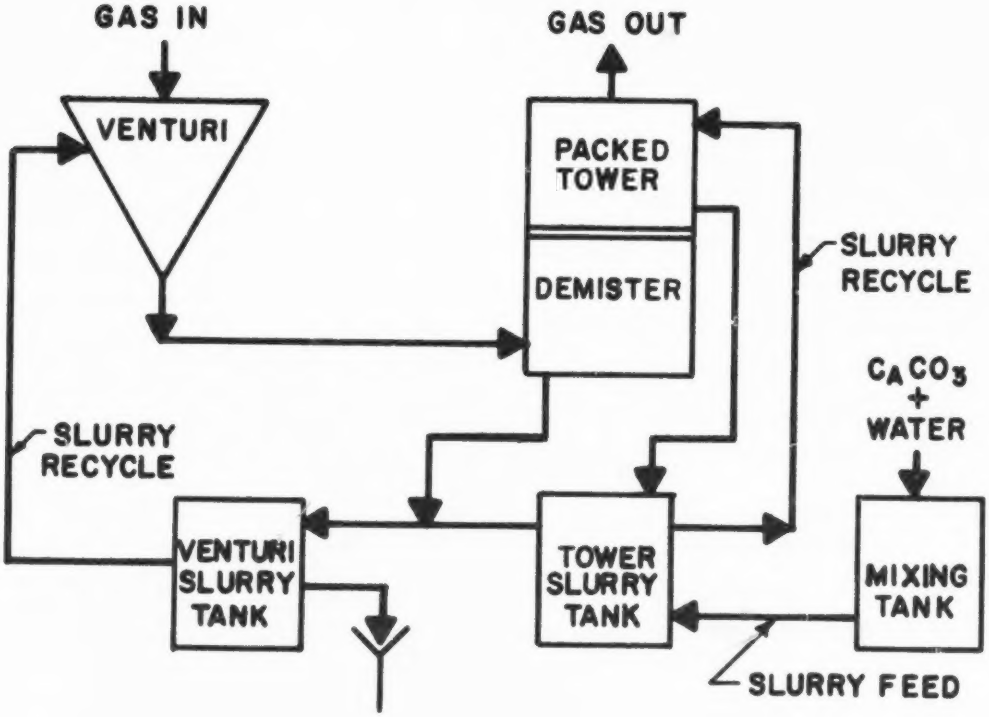
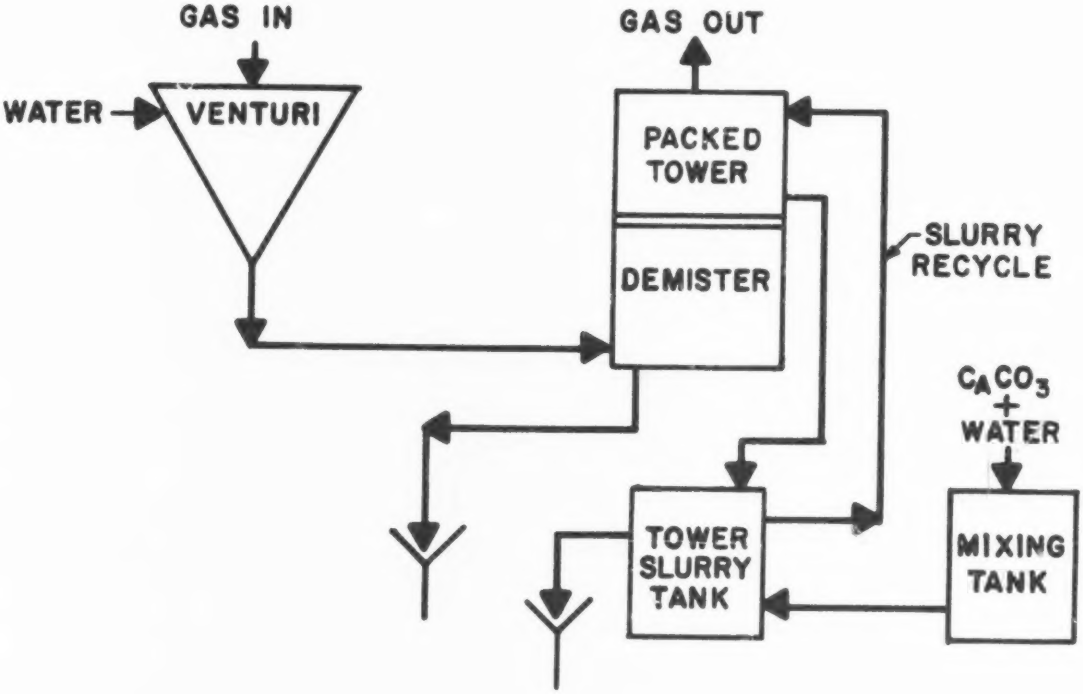
* Calcium flow rates into and out of the tower slurry tank were equal when balanced conditions prevailed.


GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
SLURRY RECYCLE
$CaCO_3$ + WATER
SLURRY RECYCLE
VENTURI SLURRY TANK
TOWER SLURRY TANK
MIXING TANK
SLURRY FEED

COUNTER-CURRENT SLURRY FLOW


GAS IN
GAS OUT
WATER
VENTURI
PACKED TOWER
DEMISTER
SLURRY RECYCLE
$CaCO_3$ + WATER
TOWER SLURRY TANK
MIXING TANK

WATER FEED TO FDS

FIG. IV-21 OPERATING MODES USED FOR LIMESTONE EFFICIENCY TESTS

76% of the inlet SO_2. Only a small fraction of the absorbed SO_2 was removed in the FDS; between 9 and 21% was sorbed with the residual alkali from the tower. The planned operating conditions for each test are listed in Table B-2, Appendix B. Detailed operating results for these experiments are given in Table IV-9; each efficiency measurement listed is an average of four readings over a three hour period.

The coarse grind limestone, 75% - 200 mesh, allowed an absorption efficiency of 88.4% while a finely ground material, 89% - 325 mesh, achieved 96% removal.

Liquid-to-gas ratio in the tower had considerable influence on SO_2 reduction. At L/G of 40 gallons per 1,000 cf (Task A6 - coarse grind limestone) the efficiency across the tower was 81.6%; however, at L/G of 20 for the same operating mode, the SO_2 removal dropped to 58.2%. Such sensitivity to the liquid flow implies a significant liquid-phase absorption resistance.

2. Limestone Scaling Experiments - Closed Loop

To determine the operating conditions for minimum scaling using limestone alkali, four continuous 40 hour tests were made. Operating parameters, such as liquid-to-gas ratio, tower slurry tank residence time, and tower slurry tank temperature were varied. The scale deposition for each test was measured by weighing the packing before and after each run; in most cases new packing was installed for the subsequent test. Following these preliminary experiments, an eighty-hour continuous operation was carried out and the scale accumulation measured. A description of the program plan is given in Table B-3.

TABLE IV-9

SUMMARY DATA SHEET FOR THE TVA TEST PROGRAM[1]

Material Used	Selma Chalk	Cement Dust	Limestone					Lime Hydrate	Limestone[2]	
			Fine	Coarse	Coarse	Coarse	Coarse		Fine	Coarse
Task No.	A2	A3	A4	A5	A6	A7	A8	A9	A4'	A5'
Gas Flow, CFM[3]	900	900	600	600	800	900	800	600	900	500
Tower Liquid Rate, GPM	35	36	24	24	32	18	32	24	36	23.7
FDS Liquid Rate, GPM	9.4	9	6	6	8	9	8	6	9	5.3
Gas Velocity, Tower, Ft/Sec.	10.7	10.7	9.3	9.3	9.6	10.7	9.6	9.3	10.7	6.0
Gas Velocity, FDS, Ft/Sec.	99	105	132	132	121	98	87	132	98	105
Tower Pressure Drop, inches H_2O	3.9	6.5	1.1	0.9	1.2	1.3	1.3	1.9	15.8	0.32
FDS Pressure Drop, inches H_2O	12.7	6.4	7.3	7.3	7.1	7.9	9.4	7.2	6.5	6.9
CaO/SO_2 Ratio	1.11	1.06	1.16	1.00	1.00	1.10	1.62	1.04	1.01	1.54
SO_2 Concentrations, PPM										
FDS in	1550	1135	1650	1980	1467	1637	1485	1544	1270	1475
FDS out	1405	1025	1288	1795	1155	1415	1402	1212	1300	1427
Tower out	48	268	62	209	212	592	199	16	193	100
Fraction of SO_2 Removed, %										
FDS	9.4	9.7	21.9	9.4	21.2[4]	13.6[4]	5.6	21.5	--	3.3
Tower	96.6	73.5	95.1	88.4	81.6	58.2	85.8	98.7	85.1	93.0
Overall	96.9	76.4	96.1	89.5	85.5	63.8	86.6	98.9	85.1	93.2
Gas Temperature, °F										
FDS in	371	377	361	361	348	391	369	357	360	355
FDS out	122	122	118	116	92	102	101	118	121	109
Tower out	114	116	110	109	91	99	98	111	113	98
Liquid Temperature, °F										
FDS in	119	121	118	116	40	40	41	121	121	106
FDS out	126	124	125	124	75	97	95	123	125	116
Tower in	112	114	110	110	89	95	96	112	115	100
Tower out	122	122	117	116	124	106	101	118	120	106
L/G ratio, FDS, Gal. per 1000 cf	10.4	10	10	10	10	10	10	10	10	10.6
L/G ratio, Tower, Gal. per 1000 cf	39	40	40	40	40	20	40	40	40	47.5

1. Operating conditions for each task shown are an average of four readings measured over a three hour period.

2. Tasks A4' and A5' were not considered at steady state condition or at the specified operating level.

3. Gas flow at tower outlet temperature and approximately 380 inches of H_2O

4. Water was fed to the FDS for this test.

The operating modes for this test series are shown in Figures IV-22 and IV-23. Originally, a closed-loop scheme, as illustrated in Figure IV-23, was proposed for all tests. Although operational problems encountered during the program required plan modifications, the continuous eighty hour run was performed with the closed-loop process. One experiment, (Task C5) as described below, was executed in a open-loop system for part of the run.

a. Scaling

The detailed operating conditions for each test are summarized in Table B-4. The preliminary experiments were 40 hour runs performed as a guide in determining the scale buildup at various operating conditions. Although the process parameters for these tasks were not constant during any run, a general trend in the scale accumulation could be seen. With high tower L/G, a lower solids buildup was measured than with low L/G, see Table IV-10, Tasks C2 and C3. Residence time for the slurry in the tower hold tank showed no effect as the hold time was varied from 5 to 10 minutes.

The profile of solids buildup on the packing looking from the top to the bottom section did show a pattern of scaling; very little scale deposited in the top section and a consistent quantity precipitated on the bottom three or four elements. This profile of solids buildup suggested an absorption-supersaturation taking place within the tower with an induced encrustation after one or two feet.

From the profile of solids deposited on the packing and the reduced deposition at high L/G, one could conclude that incoming solution from the slurry tank was at low supersaturation but once supersaturation did develop within the tower, the rate of encrustation was consistent.

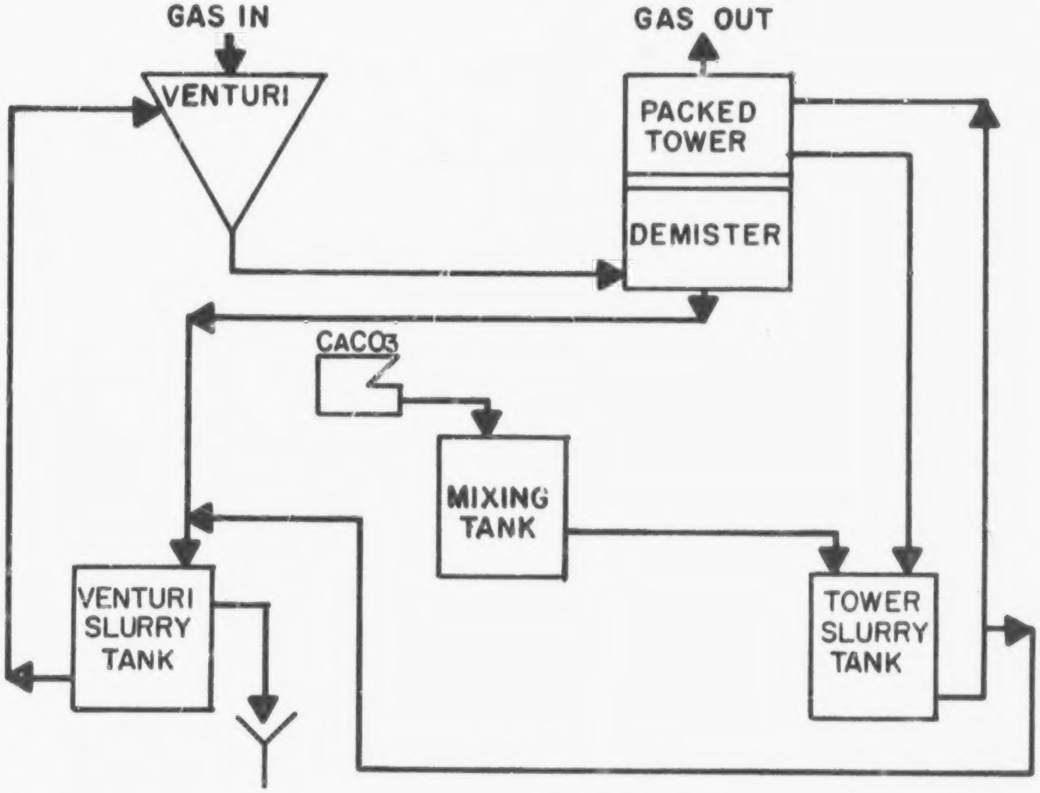

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CACO3
MIXING TANK
VENTURI SLURRY TANK
TOWER SLURRY TANK

FIG. IV-22 FLOW DIAGRAM FOR OPEN-LOOP SCALING TESTS

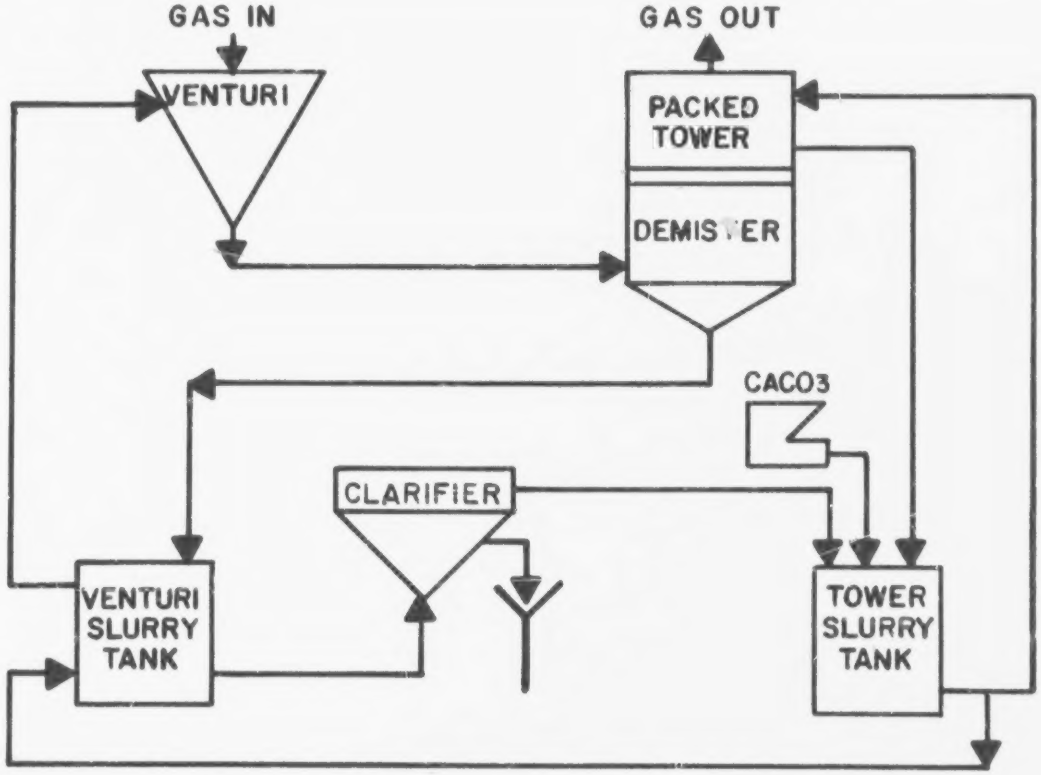

GAS IN
GAS OUT
VENTURI
PACKED TOWER
DEMISTER
CACO3
CLARIFIER
VENTURI SLURRY TANK
TOWER SLURRY TANK

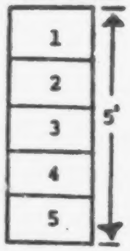
FIG. IV-23 TWO STAGE CALCIUM CARBONATE SCRUBBER

TABLE IV-10

PACKING WEIGHTS BEFORE AND AFTER EACH TASK

	Task C-2		Task C-3		Task C-4		Task C-5		Task C-6			
									(40 HOURS) *		(80 HOURS)	
Packing No.	Before lbs.	Weight Gain lbs.	Before lbs.	Weight Gain lbs.	Before lbs.	Weight Gain lbs.	Before lbs.	Weight Gain lbs.	Before lbs.	Weight Gain lbs.	Before lbs.	Weight Gain lbs.
1	8	7	4	2	8	4	22*	2	9	.5**	9	3
2	6	14	5	9	4	13	21	5	8	1.5	8	3
3	8	13	6	10	6	15	17	8	8	1.5	8	7
4	6	16	6	10	8	14	12	8	7	3.5	7	6
5	6	14	6	12	6	20	8	8	7	3.5	7	4
Total :		64		43		66		31		10.5		23

Packing Position



* Packing was weighed slightly wet.

** Weight gain was calculated by subtracting 0.5 lbs. of moisture from each section. Five new sections of packing showed 2.5 lbs. of gain when wetted with water.

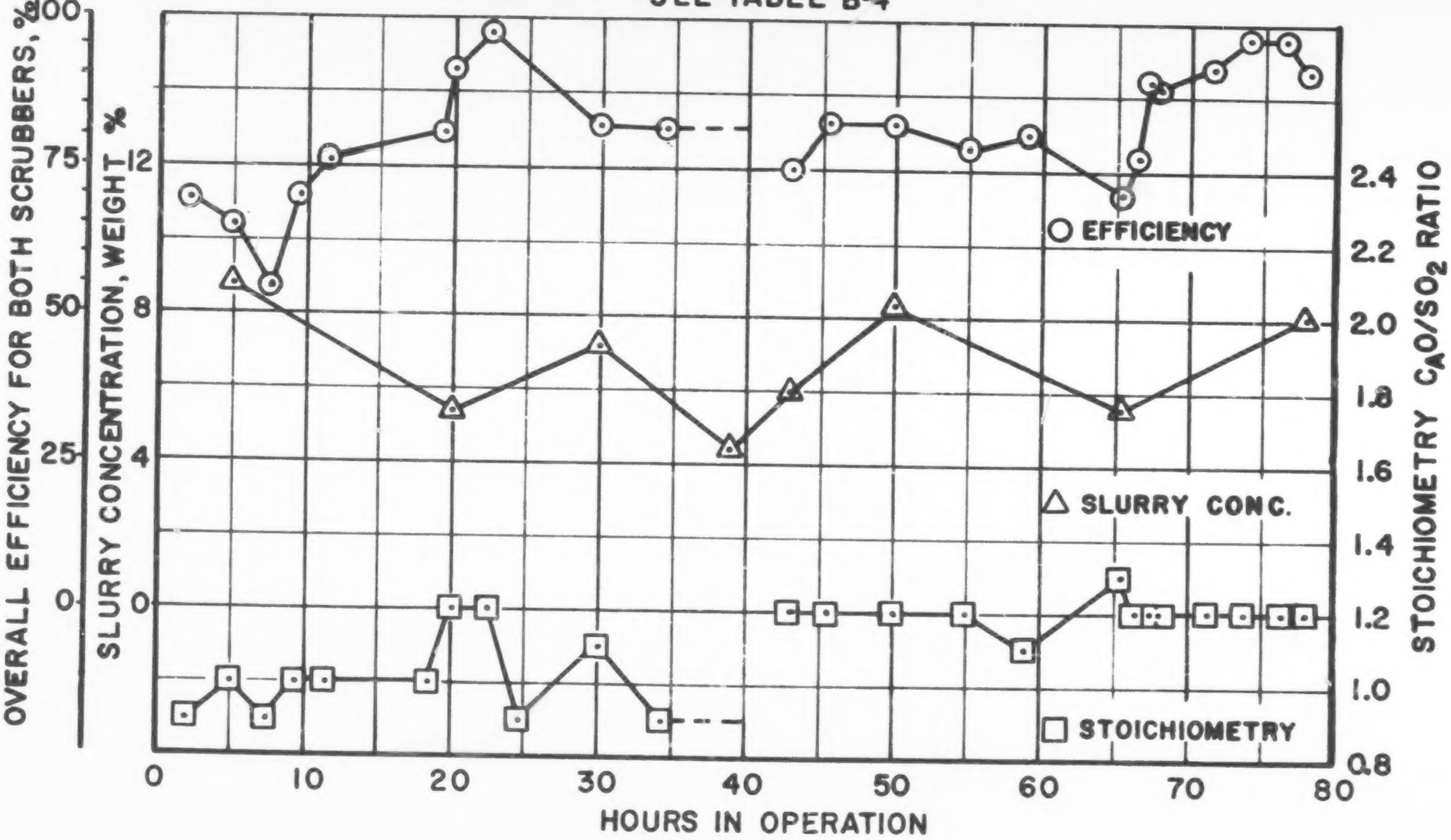
Based on the aforementioned reasoning, the operating conditions selected for the long term demonstration test combined a low level tower hold tank volume with a maximum practical tower liquid-to-gas ratio, i.e. a ten minute residence time on the hold tank and a L/G of 45 gallons per 1,000 cf.

Some time after the actual test program, a chemical analyses of the solutions entering and leaving the tower showed that the dissolved $CaSO_4.2H_2O$ was approximately the same for the tower inlet and outlet.[1] On the other hand, calcium sulfite in the solution leaving the tower was supersaturated to approximately six times its solubility; yet the liquid entering the tower (leaving the hold tank) was not supersaturated at all. Hence, the assumption of low supersaturation for the solutions leaving the hold tank was correct and the decision for high liquid-to-gas flow would tend to reduce scaling.

Stoichiometry at the start of the run was near 100% for the first 40 hours and 120% for the last 40 hours. Two limestone grinds were used during the run--for the first 40 hours a limestone having 75% - 200 mesh was employed, while during the second 40 hours, the same material with 61% - 200 mesh was used. A chemical and particle analyses for these materials are given in Table B-5. Solids concentration in the slurry was held between 4.4 and 8.9%.

The absorption efficiency varied from a low of 55% to a high of 98%. Near the end of the run, the absorption was highest. A profile plot of stoichiometry, slurry concentration and tower efficiency is shown in Figure IV-24. An explanation for the variation in efficiency is discussed in the following absorption section.

FIG. IV-24
TASK C-6 EFFICIENCY PROFILE, SLURRY CONCENTRATION
AND STOICHIOMETRY DURING THE RUN
SEE TABLE B-4



During the continuous scaling run, an equipment breakdown interrupted the test about half way through the run. The packing was removed from the tower and each section was examined and weighed. Most of the encrustation deposited during the 40 hours was on the packing periphery. The encrustation had a mud-like consistency and not the hard scale observed during the calcium oxide tests. Scale was not evident on the well-irrigated surfaces. For the second 40 hours the same packing was used; encrustation that did develop was again predominantly at the periphery. The measured weight gain after 40 and 80 hours is given in Table IV-10.

After the first 40 hours, 10.5 pounds of solids had built up, and during the next 40 hours, an additional 12.5 pounds were deposited, weight gain determined on dry basis. No pressure increase could be measured throughout the 80 hour test. The amount of solids clinging to the packing was a small fraction of the packing void volume; pressure measurement on the tower was approximately 1.0 inches of H_2O at the start and finish of the test, as given in Table B-4.

b. Absorption

The absorption efficiency for the preliminary scaling runs varied considerably. Limestone stoichiometry, slurry concentration and the inlet gas composition were changing throughout the test series. To explain the efficiency variation, the operating conditions for the 80 hour run were examined carefully. Slurry chemical analysis performed by TVA and Radian Corporation were combined with the absorption efficiency, inlet SO_2 concentration and slurry concentration measured in the field.[1] A list of the TVA analysis for test C-2 to C-6 is given in Table B-6.

By digital computer simulation of the absorption process, the calcium carbonate concentration was determined for the absorbate over the entire 80 hour run. With a known chemical analysis as a starting point, the $CaCO_3$ slurry concentration was calculated and for each point in time an efficiency measurement was made. The predicted carbonate slurry concentration fit well with the chemical analyses. The computer simulation could not take into consideration process changes such as spills, leaks or uncontrolled water addition. A comparison of the computed and analyzed carbonate concentration is given in Table IV-11.

TABLE IV-11

LIMESTONE CONCENTRATION IN THE HOLD TANK DURING TASK C6

		Calcium Carbonate Concentration, %		
Date	Time	Computer Predicted	TVA	Radian
1/25	2230	1.89 (Start)	1.89	--
1/26	1300	1.19	0.54	--
1/26	1500	1.19	--	1.13
1/26	2100	1.46	0.94	--
1/29	1400	0.36 (Start)	0.36	--
1/29	2100	1.44	2.16	--
1/30	1230	1.32	1.65	--
1/30	0103	1.78	1.60	--

Using the computer-estimated value of limestone concentration, the analyzed sulfur dioxide absorption efficiency in the tower, and the measured tower hold tank slurry concentration, a correlation was developed which predicts the absorption efficiency for each of the limestone materials employed.

For the first 40 hours, where limestone ground to 75% - 200 mesh was used, an outstandingly good correlation was realized. The absorption efficiency, predicted to within ±1.9%, showed sensitivity to inlet SO_2 concentration and limestone concentration as seen below:

$$Y = 165.05 - 0.0463\,(\text{ppm}) + 30.48\,(\%\ CaCO_3) - 9.126\,(SL) \qquad \text{(IV-9)}$$

where Y = SO_2 absorption efficiency, %,
ppm = tower inlet SO_2 concentration in ppm,
% $CaCO_3$ = concentration of limestone slurry in the hold tank, %,
SL = concentration of all solid in hold tank, %.

Statistical parameters and the variable range for this correlation are listed in Table IV-12.

For the second half of the run, a similar linear correlation having a precision of ± 3% efficiency was developed for limestone with 61% - 200 mesh. Here the last 26 hours of operation were studied so that a mixture of the two limestone types could be avoided, i.e. 14 hours of the run time were deleted because of the limestone mixture. The predicted efficiency showed less sensitivity to inlet SO_2 concentration and greater sensitivity to the limestone concentration.

TABLE IV-12

STATISTICAL PARAMETERS FOR EFFICIENCY CORRELATION

(First 40 Hours Run*)

Equation IV-9

Number of Data Points	=	30
Correlation Coefficient	=	0.987
Standard Error For Estimate	=	1.9
Significance of Regression (F)	=	349
% Efficiency Range, Y	=	53% to 97%
Sulfur Dioxide Inlet Range	=	1160 to 1900 ppm
% $CaCO_3$ Range	=	1.131% to 1.89%
% Total Solids Range, SL	=	5.4% to 9%

---

* Limestone Used - Tiftona Limestone 50.5% CaO, 75% - 200 Mesh.

$$Y = 56.273 - 0.0178\,(\text{ppm}) + 50.313\,(\%\ CaCO_3) - 4.15\,(SL) \qquad \text{(IV-10)}$$

See Table IV-13 for statistical limitations.

To make use of these efficiency correlations, the limestone slurry concentrations must be known. Three factors influence the residual limestone concentration in the tower slurry liquor: 1) the actual absorption efficiency for the process, 2) the stoichiometric feed ratio of $CaCO_3/SO_2$ and 3) the overall slurry concentration. For a system with 6% total slurry, the limestone concentration can be predicted by:

$$\%\ CaCO_3 = 1.23 - 0.033\,(\%\ \text{eff.}) + 2.236\,R \qquad \text{(IV-11)}$$

where R = stoichiometric feed ratio, mols of $CaCO_3$/mole SO_2.

Table IV-14 presents the conditions for equation (IV-11).

Using this expression and equation (IV-9) or (IV-10), the absorption efficiency can be predicted for either limestone material for a liquid-to-gas ratio of 45 gallons per 1,000 cf., and a total slurry of 6% by weight.

Clearly, if the absorption efficiency is dependent upon the limestone slurry concentration and the inlet gas SO_2 concentration, then one or both of these conditions must be controlled for a desired SO_2 removal. Reviewing once again the 80-hour demonstration run, the computer-predicted carbonate slurry concentration was compared to the material balance expression, equation (IV-11).

TABLE IV-13

STATISTICAL PARAMETERS FOR EFFICIENCY CORRELATION

(Last 26 Hours Run*)

Equation IV-10

Number of Data Points = 26

Correlation Coefficient = 0.95

Standard Error For Estimate = 3.08

Significance of Regression (F) = 69

% Efficiency Range, Y = 75.5% to 97.9%

Sulfur Dioxide Inlet Range = 960 to 1380 ppm

% $CaCO_3$ Range = 1.317% to 1.847%

% Total Solids Range, SL = 5.5% to 8%

---

* Limestone Used - Tiftona Limestone 50.8% CaO, 61% - 200 Mesh.

TABLE IV-14

STATISTICAL PARAMETERS FOR
STEADY STATE WT.% LIMESTONE CORRELATION

Number of Data Points = 16

Correlation Coefficient = 0.995

Standard Error For Estimate = 0.054

Significance of Regression (F) = 741.8

% $CaCO_3$ Range = 0.1616% to 2.07%

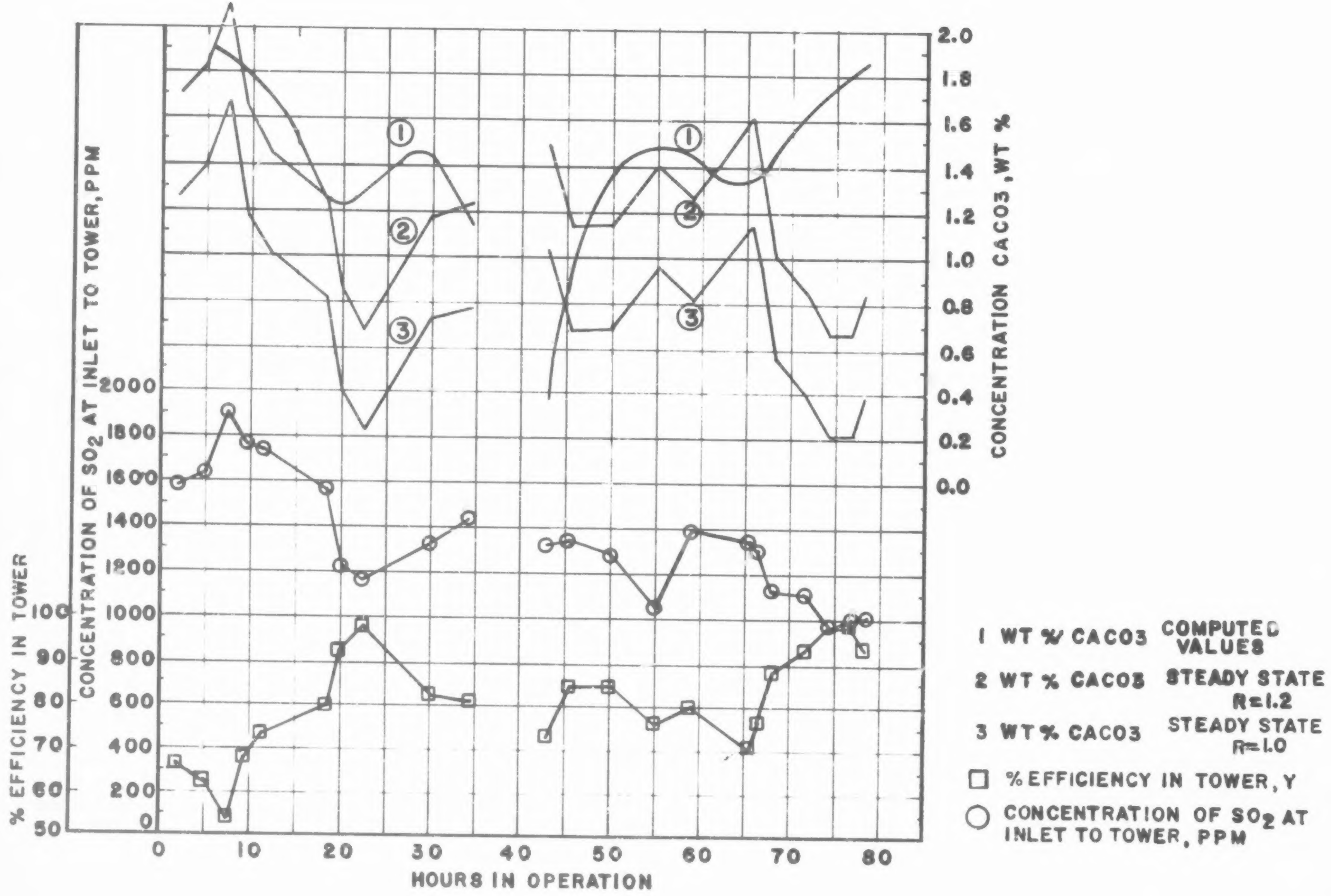
% Efficiency Range, Y = 53% to 92%

Range For Stoichiometric Ratio = 1.0 to 1.3

At several points during the test, the limestone steady-state composition at R = 1.20 and the predicted limestone concentration were coincident, as illustrated in Figure IV-25. For these points of intersection, all conditions for equation (IV-11) and (IV-9) or (IV-10) could be satisfied by a limestone stoichiometry of 120%.

The computed values for calcium carbonate stoichiometric ratio for R = 1.0 was included in Figure IV-25 to illustrate the change in calcium carbonate concentration. Equations (IV-9) through (IV-11) should be most useful in planning the Shawnee test program for direct limestone addition to the scrubbing circuit.

FIG. IV-25 PROCESS CONDITIONS FOR TASK C6



V. CONCLUSIONS

1. Sulfur dioxide absorption with sodium carbonate in a FDS contactor can be varied by controlling the pressure drop across the venturi throat. Other operating parameters, i.e. liquid-to-gas ratio, stoichiometric ratio or SO_2 concentration did not significantly affect performance for the range of conditions tested.

2. Calcium oxide absorption of SO_2 is less efficient than Na_2CO_3 for similar operating conditions. Variations in liquid-to-gas ratio, stoichiometry, throat pressure drop, slurry concentration and ionic strength all affected the SO_2 absorption efficiency. Sensitivity of these variables indicates a significant liquid phase mass-transfer resistence.

3. With lime reagent, the scale formation is rapid and severe. High slurry concentration (15 to 20% by weight) through the venturi scrubber did not eliminate the severe encrustation of the FDS internals.

4. Addition of sodium chloride to a slurry of calcium hydroxide improves the sulfur dioxide absorption. The increase in ionic strength with the NaCl simulates, to some extent, the steady-state conditions for a closed-loop system.

5. Sulfated lime/fly ash material from the Shawnee Power Station showed significantly lower absorption efficiency than the calcium oxide at comparable operating conditions.

6. Dolomitic lime (CaO.MgO) demonstrated outstandingly good absorption efficiency for the single test made.

7. A limestone slurry, circulating through a high specific-surface packed tower can absorb greater than 90% of the flue gas SO_2. Limestone utilization in the experimental tower was 78% at 1000 ppm SO_2 and 60% at 1500 ppm SO_2. Absorption efficiency is adversely affected by increasing SO_2 concentration and by high slurry concentration. SO_2 absorption can be improved by increasing the calcium carbonate slurry concentration in the absorbing liquor.

8. A finely ground limestone (90% - 325 mesh) increases the SO_2 absorption by 8 to 10% over a material with (75% - 200 mesh). Absorption increases with higher liquor-to-gas ratio in the tower.

9. Scale formation in a limestone/SO_2 scrubbing system can be controlled by maintaining a reaction product slurry in the absorbing liquor and by circulating a high liquid flow

rate through the tower. Stagnant non-irrigated areas should be avoided in the absorber design.

10. The demonstrated ability of the limestone system to remove SO_2 to low levels and the short term significant reduction in scaling behavior experienced in the present limestone tests indicate the commercial applicability of the system.

VI. RECOMMENDATIONS

1. Results of the limestone studies indicate a strong influence of SO_2 concentration and slurry composition on efficiency. Further work in this area is needed to define these effects over a broader range of conditions.

2. Composition of the alkaline solution has a striking effect on the SO_2 absorption as was demonstrated in the experiments varying the ionic strength and slurry concentrations. Future studies with lime or limestone should include, as part of the program, thorough analyses of the liquor phase.

3. Dolomitic lime demonstrated outstandingly high absorption efficiency. This material should be tested in depth in future limestone studies.

4. Scaling of the pilot unit with lime slurries was severe throughout this program. Future studies with lime injection should be considered with controlled pH.

5. The high SO_2 removal efficiency obtained with limestone plus the promising reduction in scaling obtained call for a major program devoted to exploration and exploitation of these results.

VII. REFERENCES

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APPENDIX A

OPERATING CONDITIONS AND RESULTS
For
THE SODIUM CARBONATE AND CALCIUM OXIDE TESTS

TABLE A-1

CHEMICAL AND PHYSICAL ANALYSIS OF CALCINED LIMESTONE

Particle Size	100%-200 Mesh	
Loss on Ignition	3.15%	Loss Free Basis
CaO	94.40	97.47
MgO	0.76	0.78
Al_2O_3	0.45	0.46
Fe_2O_3	0.11	0.11
SiO_2	0.96	0.90
CO_2	0.8	--

TABLE A-2

PILOT TEST PLAN

TASK I - Pilot Plant Modification and Calibration

IA - Engineering and Purchasing of additional components, i.e., hold tank, mix tank, piping, agitators, pumps, etc.

IB - Install hold tank

IC - Install mix tank

ID - Piping modifications and additions to allow for all modes.

IE - Install agitators

IF - Install and calibrate venturi flowmeters for both gas and liquid flow measurements.

IG - Install and calibrate analytical equipment including SO_2 analyzer, NO_x analyzer, temperature recorder, pH meter, etc.

TASK II - Utilization of Sodium Carbonate Slurry For Determination of Optimum Operating Conditions and Maximum Efficiency of Scrubber

A - Operating as shown in Figure IV-1, 5 levels of Δp, at 5 levels of gas flow will be tested. (25 tests)

B - Operating as shown in Figure IV-1, tests will be conducted at four levels of liquid-to-gas ratio, at two levels of gas flow. (8 tests).

C - Operating as shown in Figure IV-2, fresh water will be fed to the venturi and carbonate slurry to the packed tower at 3 levels of L/G, at 3 levels of Δp and 2 levels of gas flow. (18 tests)

TASK III - Evaluation of the Contribution of Calcined Limestone to SO_2 Removal During Capture in the Venturi

III(a) - Introduce dry calcined limestone to the gas stream at four stoichiometric levels using standard predetermined operating conditions and measure SO_2 removal. Mode of operation is shown in Figure IV-5. (4 tests).

TABLE A-2 cont'd

TASK IV - Measurement of the SO_2 Removal When a Lime Slurry is Introduced to the Venturi

IV(a) - As is shown in Figure IV-7, a lime slurry is fed to the venturi on a once through basis at four concentrations. Measure pH at venturi sump and hold tank. (4 tests)

TASK V - Measurement of SO_2 Removal Efficiency When Venturi and Packed Bed are Operated in Series

V(a) - As shown in Figure IV-8, lime is fed to hold tank; the venturi and packed tower are operated in series and solids are accumulated in the clarifier. For the standard operating conditions and when steady-state has been achieved, SO_2, pH and temperature measurements will be made at all points shown in Figure III-1. (1 test)

TASK VI - The Effect of Major Process Variables Will Be Studied for The Integrated Venturi-Packed Bed System Where the Additive is Calcined Limestone

VI(a) - Inject dry additive into gas stream as per Figure IV-9. Vary lime stoichiometry at 4 levels. All other parameters held constant at standard levels. (4 tests)

VI(b) - With mode as per Figure IV-10 vary L/G ratio. Hold liquid hold time constant by varying level in hold tank. (3 tests).

VI(c) - Vary Δp of venturi at 2 L/G ratios, all other parameters at constant standard levels. Mode as per Figure IV-10. (6 tests)

VI(d) - Vary hold time at constant L/G and constant Δp. Mode as per Figure IV-10. (3 tests)

VI(e) - Vary slurry concentration (3 levels) to venturi by dilution of liquid to venturi with clarifier overflow as per Figure IV-11. All parameters held constant except slurry concentration. (3 tests).

VI(f) - Vary ionic strength (3 levels) of scrubbing liquor by salt addition. Operate as per mode in Figure IV-10. (3 tests).

VI(g) - Determine power requirements for operating conditions described in VII-a through VII-f from the data obtained.

TABLE A-2 cont'd

TASK VII - Investigate the Effect of Mode Change

VII(a) - Add the dry lime to the hold tank and feed the packed bed recycle directly to the clarifier as per Figure IV-13. Holding all parameters constant, measure SO_2 removal. (1 test)

VII(b) - Split clarifier overflow and underflow between packed bed and venturi as per Figure IV-15. Measure SO_2 removal. (1 test)

VII(c) - Deliver clarifier overflow to venturi as shown in Figure IV-16. (1 test)

TASK VIII - Investigate the Effect of Changing the Nature of the Additive

VIII(c) - Inject calcined dolometic limestone into the gas stream as shown in Figure IV-17. Operate at standard constant conditions and determine SO_2 removal efficiencies. (1 test)

VIII(d) - Conduct one run with partially sulfated lime/fly ash mixture feeding the dry additive in the inlet stream as per Figure IV-17. (1 test)

VIII(e) - Conduct one run with partially sulfated lime/fly ash mixture feeding the lime/fly ash to the hold tank (with at least one hour hold time) as per Figure IV-19. (1 test)

TASK IX - Data Reduction and Analysis

IX(a) - Data analysis

IX(b) - Tabulation and graphical representations

IX(c) - Work session with NAPCA

TASK X - Final Report Preparation and Presentation

X(a) - Prepare draft of final report

TABLE A-2 cont'd

X(b) - Prepare graphics

X(c) - Work session with NAPCA

X(d) - Final draft presentation

X(e) - Reproduction and binding - 200 copies

X(f) - Presentation

PROGRAM MANAGEMENT:

Monthly Reporting

Contract Administration

TABLE A-3

TEST RESULTS USED IN FDS SODIUM CARBONATE CORRELATION

Test No.	Date	Time	Inlet SO_2 Concentration ppm	Outlet SO_2 Concentration ppm	ΔP In. W.G.	Gas Flow SCFM	L/G	Stoichiometric Ratio Na_2CO_3/SO_2	Liquid Rate gpm	Gas Flow Inlet SCFM	Disc Position	Throat Velocity Ft./Sec.
1	9/23	17:30	1640	920	3.0	395	12.9	3.5	5	406	1	51
2	9/23	18:00	1870	980	4.6	395	7.8	3.0	5	508	2	111
3	9/23	18:30	1970	830	2.9	395	7.8	2.88	5	508	3	162
4	9/24	16:00	1240	375	8.0	719	7.8	4.32	9	916	0	127
5	9/24	16:30	1260	260	11.0	719	7.5	4.33	9	916	1	160
6	9/24	17:10	1170	374	8.0	719	7.6	2.36	9	920	0	126
7	9/24	17:20	1170	296	11.2	719	7.8	2.36	9	920	1	159
8	9/29	17:00	1750	798	4.3	569	9.6	2.07	8	720	0	96
9	9/29	17:45	1970	865	5.7	569	10.0	1.84	9	707	1	123
10	9/29	18:00	2030	750	9.3	569	9.7	2.03	9	701	2	163
11	9/30	16:30	1370	890	1.90	246	9.7	4.70	4	300	3	97
12	9/30	16:55	1540	655	5.3	246	9.7	4.15	4	300	3	97
13	10/01	8:50	1690	940	1.85	395	8.0	4.34	5	482	0	66
14	10/01	9:40	1640	940	2.35	395	8.0	4.47	5	482	1	84
15	10/01	9:50	1620	940	4.15	438	7.2	4.08	5	482	2	124
16	10/01	10:10	1640	940	11.85	395	8.0	4.43	5	482	3	163
17	10/01	11:15	1550	845	10.75	395	8.0	4.15	5	482	3	163
18	10/01	11:30	1450	940	0.25	395	8.0	4.43	5	482	2	111
19	10/01	11:40	1500	940	2.15	395	8.0	4.29	5	482	1	84
20	10/01	11:50	1400	940	1.65	395	8.0	4.59	5	482	0	66
21	10/05	17:25	2220	890	4.53	622	9.5	3.97	9	681	0	106
22	10/05	18:25	2350	470	16.23	557	10.5	4.27	9	681	2.5	190
23	10/05	18:37	2160	515	16.23	557	10.6	4.65	9	681	2.5	190
24	10/06	10:45	1870	704	5.80	728	6.7	3.84	8	816	0	128
25	10/06	11:14	1920	680	8.0	728	6.7	3.74	8	816	1	161
26	10/06	11:35	2060	610	14.6	645	7.6	3.94	8	816	2	190
27	10/06	11:45	2060	470	10.0	645	7.6	4.27	8	816	1.5	163
28	10/06	12:00	2010	610	10.0	645	7.6	4.38	8	816	1.5	163
29	10/08	15:50	1000	645	1.65	443	6.8	4.24	5	505	0	77
30	10/08	16:05	1020	625	2.30	399	7.7	4.54	5	504	1	88
31	10/08	16:35	1000	400	14.45	399	7.7	4.61	5	504	3	171
32	10/08	16:55	1000	175	28.75	399	7.8	4.59	5	501	3.75	254
33	10/08	18:06	1000	350	12.25	719	7.5	4.49	9	908	0	128
34	10/08	17:55	1000	324	12.25	719	7.6	4.57	9	908	0.5	143
35	10/08	17:30	1000	275	15.15	802	6.8	4.10	9	908	1.5	206
36	10/08	17:20	1000	250	20.05	719	7.6	4.47	9	904	2	215
37	10/06	10:10	1830	470	14.15	449	13.7	3.78	10	492	3	192
38	10/06	15:30	1830	515	14.15	449	13.7	3.78	10	492	3	192
39	10/07	14:15	1920	515	14.40	524	7.1	5.23	6	589	2.80	205
40	10/07	15:05	1870	515	14.40	524	7.1	5.37	6	589	2.80	205
41	10/07	15:05	1750	515	14.40	521	9.4	5.77	8	585	2.80	203
42	10/07	15:25	1700	470	14.40	521	9.4	6.02	8	585	2.80	203
43	10/07	16:30	1110	236	14.40	524	11.8	5.35	10	573	2.70	197
44	10/07	16:40	1080	234	14.40	524	11.8	5.50	10	573	2.70	197
45	10/07	16:50	1110	234	14.40	524	14.1	5.35	12	573	2.5	182
46	10/07	17:00	1080	210	14.40	524	14.1	5.50	12	573	2.5	182

1. Velocities based on inlet gas conditions

TABLE A-4

OPERATING CONDITIONS FOR TASK III*

Date	10/21	10/21	10/21	10/21	10/21	10/21	10/21	10/21	10/21	10/21	10/21
Time	13:55	13:55	14:47	15:15	15:45	11:05	11:30	11:45	10:00	10:25	10:40
Gas Flow, cfm	1021	1021	1021	1021	1021	1027	1024	1024	1037	1029	1024
FDS L/G ratio, gal/mcf	6.8	6.8	7.0	6.9	6.9	7.0	6.5	6.8	6.6	6.5	-
Tower L/G, ratio, gal/mcf	14.2	14.2	14.2	14.2	14.2	7.1	7.1	7.1	7.1	7.1	-
Gas Velocity FDS, ft/sec	181	152	152	181	181	184	184	184	186	184	184
Tower pressure drop, inches H_2O	0.7	0.7	0.7	0.7	0.7	0.6	0.7	0.7	0.7	0.7	0.7
FDS pressure drop, inches H_2O^2	10.0	10.0	10.0	10.0	10.0	10.0	10.0	9.8	10.0	10.0	10.0
CaO/SO_2 ratio	2.25	2.52	1.86	1.99	2.13	1.53	1.47	1.53	0.97	2.75	0.98
SO_2 concentration, ppm											
FDS in	640	605	640	622	675	875	807	775	972	-	960
FDS out	337	337	404	320	404	505	505	470	672	-	640
Tower out	202	202	202	236	0	278	135	218	219	-	118
Fraction of SO_2 removed, %											
FDS	47.3	44.3	36.9	48.6	40.2	42.3	37.4	39.4	30.9	-	33.3
Tower	40.1	40.1	50.0	26.3	100.0	45.0	73.3	53.6	67.4	-	81.6
Overall	68.5	66.6	68.4	62.0	100.0	68.2	83.3	71.8	77.4	-	87.7
Gas Temperatures, °F.											
FDS in	338	338	338	338	338	340	338	338	348	342	338
FDS out	105	105	105	105	105	105	105	105	105	108	105
Tower out	72	72	70	70	70	99	88	88	88	88	88
Liquid Temperatures, °F.											
FDS in	62	62	64	62	62	62	62	62	62	62	62
FDS out	98	98	98	98	98	98	98	98	98	98	98
Tower in	-	-	-	-	-	-	-	-	-	-	-
Tower out	95	92	92	92	92	100	100	100	100	100	100
pH Measurements											
Tower outlet	-	-	-	-	-	-	-	-	-	-	-
FDS outlet	11.4	11.4	11.0	10.8	10.8	10.7	11.3	11.3	6.5	7.6	8.2
Hold Tank	-	-	-	-	-	-	-	-	-	-	-
Clarifier tank	-	-	-	-	-	-	-	-	-	-	-
Disc Position	1.30	1.30	1.30	1.90	1.90	1.94	1.94	1.94	1.94	1.94	1.94
Pressure at outlet, in. Hg gauge	-1.7	-1.7	-1.8	-1.8	-1.8	-1.8	-1.8	-1.8	-1.8	-1.8	-1.8

* L/G on FDS based on inlet gas conditions.

TABLE A-4 cont'd

OPERATING CONDITIONS FOR TASK III*

Date	10/22	10/22	10/22	10/22
Time	15:20	15:50	15:50	16:15
Gas Flow, cfm	1030	1030	1030	1030
FDS L/G ratio, gal/mcf	6.7	6.7	6.7	6.7
Tower L/G, ratio, gal/mcf	14.2	14.2	14.2	14.2
Gas Velocity FDS, ft/sec	182	182	182	182
Tower pressure drop, inches H_2O	0.8	0.8	0.8	0.80
FDS pressure drop, inches H_2O	10.0	10.0	10.0	10.0
CaO/SO_2 ratio	0.80	0.80	1.15	1.04
SO_2 concentration, ppm				
FDS in	1275	1275	1260	1345
FDS out	925	925	872	975
Tower out	455	404	353	370
Fraction of SO_2 removed, %				
FDS	27.5	27.5	30.8	27.5
Tower	50.8	56.3	59.5	62.1
Overall	64.3	68.3	72.0	72.5
Gas Temperatures, °F.				
FDS in	350	350	350	360
FDS out	108	102	102	112
Tower out	72	72	72	88
Liquid Temperatures, °F.				
FDS in	62	62	62	70
FDS out	100	95	95	108
Tower in	-	-	-	-
Tower out	95	95	92	102
pH Measurements				
Tower outlet	-	-	-	-
FDS outlet	8.1	7.8	8.8	5.5
Hold tank	-	-	-	12.0
Clarifier tank	-	-	-	-
Disc Position	1.90	1.90	1.90	1.75
Pressure at outlet, in. Hg gauge	-1.8	-1.8	-1.8	-1.9

TABLE A-5

OPERATING CONDITIONS FOR TASK IV*

Date	10/27	10/27	10/27	10/27	10/27	10/27	10/27	10/27	10/27	10/27
Time	11:55	12:05	10:55	11:20	11:45	9:35	10:00	15:30	15:50	14:20
Gas Flow, cfm	1049	1049	1049	1049	1049	1047	1047	1049	1049	1047
FDS L/G ratio, gal/mcf	6.6	6.6	6.9	6.8	--	6.0	6.0	6.4	6.6	6.7
Tower L/G, gal/mcf	7.1	7.1	21.3	21.3	--	21.3	21.3	21.3	21.3	21.3
Gas Velocity FDS, ft/sec.	177	177	177	177	177	177	177	177	177	177
Tower pressure drop, inches H_2O	0.8	0.8	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
FDS pressure drop, inches H_2O	9.7	9.7	10.0	10.0	10.0	10.0	10.0	9.9	3.8	10.0
CaO/SO_2 ratio	0.78	0.78	0.75	--	--	0.96	--	1.39	1.33	1.15
SO_2 concentration, ppm										
FDS in	1580	1580	1640	1600	1600	1600	1640	1680	1760	1600
FDS out	940	940	1000	980	960	940	882	882	882	962
Tower out	600	600	710	520	450	300	225	330	330	375
Fraction of SO_2 removed, %										
FDS	40.5	40.5	39.0	38.8	40.0	41.3	46.2	47.5	49.9	39.9
Tower	36.2	36.2	26.0	46.9	52.1	68.1	74.5	62.6	62.6	61.0
Overall	62.0	62.0	54.9	67.5	71.3	81.2	86.3	80.4	81.2	76.6
Gas Temperature, °F.										
FDS in	360	360	360	360	360	358	358	360	360	358
FDS out	112	112	110	110	108	105	110	105	105	108
Tower out	88	92	68	68	68	68	58	65	65	68
Liquid Temperature, °F.										
FDS in	70	70	70	70	70	72	72	65	68	68
FDS out	108	106	108	108	108	104	108	102	102	105
Tower in	--	--	--	--	--	--	--	--	--	--
Tower out	102	105	92	92	92	88	92	88	88	92
pH measurements										
Tower outlet	--	--	--	--	--	--	--	--	--	--
FDS outlet	5.5	5.6	5.4	5.5	5.5	5.7	5.7	7.4	7.2	5.7
Hold Tank	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0
Clarifier Tank	--	--	--	--	--	--	--	--	--	--
Disc Position	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75
Pressure at outlet, in. Hg gauge	-1.8	-1.8	-2.0	-2.0	-2.0	-2.0	-2.0	-1.9	-1.9	-2.0

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* L/G on FDS based on inlet gas conditions.

TABLE A-6

OPERATING CONDITIONS FOR TASK V

Date	2/4	2/4	2/9	2/9	2/10	2/11	2/11	2/15	2/15
Time	1500	1620	1300	1700	1445	1100	1430	2000	2130
Gas Flow, cfm	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	18	18	18	18	18	18	18	18	18
Tower L/G, ratio, gal/mcf	15	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec *	118	118	78	80	93	77	77	77	77
Tower pressure drop, inches H_2O	0.4	0.4	0.4	0.4	0.4	0.4	0.3	0.4	0.4
FDS pressure drop, inches H_2O	6.0	6.0	6.0	6.0	6.5	11.4	22.0	7.0	9.1
CaO/SO_2 ratio	0	0	0.58	0.71	0.69	1.23	1.11	0.93	1.43
SO_2 concentration, ppm									
FDS in	1320	1310	1360	1080	1460	1820	1660	2000	2000
FDS out	1120	1125	124	620	830	860	700	1070	1035
Tower out	770	775	560	290	605	440	440	450	380
Fraction of SO_2 removed, %									
FDS	15	14.1	--	42.7	43.2	42.3	58	46.4	53.0
Tower	31.3	31.2	--	53.3	27.2	48.8	37	58.0	63.4
Overall	41.5	40.9	58.9	73.2	58.5	75.8	73.5	77.5	82.7
Gas temperatures, °F									
FDS in	348	346	348	352	365	370	362	--	--
FDS out	119	110	120	120	128	129	122	130	131
Tower out	90	92	112	115	115	118	118	115	121
Liquid temperatures, °F									
FDS in	105	98	118	110	121	124	112	115	118
FDS out	118	110	122	122	128	129	122	130	131
Tower in	108	98	118	114	123	125	118	118	118
Tower out	111	106	120	120	126	128	122	129	130
Clarifier	60	70	112	113	113	110	112	101	112
pH measurements									
Tower outlet	<2	<2	2.6	5.1	4.2	4.5	5.7	6.0	5.9
FDS outlet	2.4	2.0	5.6	5.1	5.4	6.4	8.8	7.6	10.1
Hold tank	4.4	4.0	11.8	11.4	11.4	11.1	11.1	10.9	10.9
Clarifier tank	4.4	3.0	5.8	5.6	5.6	6.1	10.1	11.2	10.7
Disc Position	1.5	1.5	0.1	0.2	0.8	0	0	0	0

* Gas velocity through the tower for all tests in this series was 8.6 ft/sec.

TABLE A-7

OPERATING CONDITIONS FOR TASK VI-a

Date	2/16	2/16	2/17	2/17	2/17	2/17	2/18	2/18
Time	1020	1115	1130	1430	1745	1830	1400	1730
Gas Flow, cfm	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	18	18	18	18	18	18	18	18
Tower L/G ratio, gal/mcf	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec	77	77	77	99	77	77	77	77
Tower pressure drop, inches H_2O	0.3	0.3	0.4	0.4	0.7	0.7	0.8	0.8
FDS pressure drop, inches H_2O	6.0	6.0	5.0	6.2	6.0	6.0	7.7	7.7
CaO/SO_2 ratio	0.89	0.96	1.62	1.14	1.12	1.09	0.96	0.985
SO_2 concentration, ppm								
FDS in	2280	2160	1515	1635	1680	1680	1485	1500
FDS out	2010	1320	780	875	840	660	885	900
Tower out	1600	1040	20	0	0	0	0	0
Fraction of SO_2 removed, %								
FDS	12.7	39.0	51.5	46.5	50.0	60.7	40.5	40.0
Tower	20.4	21.2	97.3	100	100	100	100	100
Overall	29.8	51.9	99	100	100	100	100	100
Gas temperatures, °F								
FDS in	--	--	--	--	--	--	--	--
FDS out	135	132	135	135	142	140	135	128
Tower out	118	118	125	129	138	138	126	124
Liquid temperatures, °F								
FDS in	118	118	120	123	144	142	135	126
FDS out	138	136	136	135	144	142	138	132
Tower in	122	122	125	128	146	146	136	128
Tower out *	72	72	81	78	80	80	136	132
Clarifier	112	112	115	125	136	136	116	118
pH Measurements								
Tower outlet	5.8	4.8	5.8	7.3	8.0	8.0	5.1	5.0
FDS outlet	12.0	12.0	10.5	10.7	11.0	11.0	11.8	10.5
Hold tank	3.7	3.7	4.2	6.3	5.8	7.0	4.4	4.2
Clarifier tank	6.0	6.2	11.6	11.6	11.6	11.6	11.1	10.7
Disc Position	0	0	0	1.0	0	0	0	0

* Tower out thermocouple was not working for low value readings.

TABLE A-8

OPERATING CONDITIONS FOR TASK VI-b

Date	2/24	2/24	2/24	2/24	2/24	2/24	2/24	2/24	2/24
Time	1100	1215	1315	1400	1500	1600	1730	1800	1830
Gas Flow, cfm	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	18	18	18	10	10	10	23	23	23
Tower L/G ratio, gal/mcf	15	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec	81	81	77	96	99	99	77	77	77
Tower pressure drop, inches H_2O	0.6	0.6	0.6	0.5	0.6	0.6	0.6	0.6	0.6
FDS pressure drop, inches H_2O^2	6.0	6.0	6.0	5.8	6.1	5.9	6.4	6.8	6.8
CaO/SO_2 ratio	1.05	1.17	1.21	-	0.93	0.81	1.06	1.01	1.07
SO_2 concentration, ppm									
FDS in	1350	1350	1410	--	1650	1800	1860	1920	2060
FDS out	720	690	840	--	1080	1200	990	1020	945
Tower out	100	188	460	--	680	800	640	780	640
Fraction of SO_2 removed, %									
FDS	46.5	49.0	40.5	--	34.5	32.5	46.8	47.0	54.3
Tower	86.2	72.6	45.3	--	37.0	33.2	35.3	23.5	32.3
Overall	92.6	86.0	68.3	--	59.0	55.5	65.6	59.4	69.0
Gas temperatures, °F									
FDS in	320	320	320	335	335	355	360	360	360
FDS out	140	150	152	155	152	152	140	150	150
Tower out	122	140	148	150	150	148	135	148	140
Liquid temperatures, °F									
FDS in	122	142	149	150	150	142	124	142	142
FDS out	138	150	152	156	155	152	138	150	150
Tower in	142	154	159	160	148	142	138	150	152
Tower out	138	150	150	153	150	150	138	144	146
Clarifier	122	138	145	152	148	140	120	135	136
pH Measurements									
Tower outlet	5.2	5.2	5.2	5.1	5.0	5.1	5.7	5.8	5.7
FDS outlet	9.8	9.6	9.3	9.5	9.4	9.4	9.8	9.8	9.8
Hold tank	10.8	10.2	10.2	-	10.2	10.8	9.8	9.6	10.0
Clarifier tank	11.2	10.7	10.4	10.3	10.4	10.4	10.6	10.6	10.6
Disc Position	0.25	0.25	0	0.9	1.0	1.0	0	0	0
Hold tank volume	470	460	460	280	250	250	650	690	590

TABLE A-9

OPERATING CONDITIONS FOR TASK VIC

Date	3/30	3/30	3/31	3/31	4/1	4/1	4/1	4/1	4/5	4/7*	4/7*	4/7*	4/7*
Time	1400	1630	1000	1200	1000	1200	1400	1700	1545	1130	1330	1600	2000
Gas Flow, cfm	700	700	700	700	700	700	700	700	680	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10.3	10	10	10	10
Tower L/G, ratio, gal/mcf	15	15	15	15	15	15	15	15	15.4	15	15	15	15
Gas Velocity FDS, ft/sec	162	138	151	130	142	124	143	181	197	277	231	194	195
Tower pressure drop, inches H_2O	0.5	0.6	0.6	0.7	0.6	0.7	0.7	0.8	0.7	0.6	0.7	0.8	0.8
FDS pressure drop, inches H_2O	6.5	6.4	6.0	6.5	6.3	6.2	6.1	12.0	12.0	12.6	12.7	11.7	13.0
CaO/SO_2 ratio	1.04	0.97	1.04	1.21	0.96	1.02	1.00	0.97	1.02	1.01	1.04	1.24	0.96
SO_2 concentration, ppm													
FDS in	1935	1860	2160	1680	1920	2040	1890	2040	1920	2220	1765	1485	1530
FDS out	960	810	1110	1080	1120	1170	1080	1200	1200	810	415	363	450
Tower out	98	360	540	510	570	600	600	760	58	360	146	116	190
Fraction of SO_2 removed, %													
FDS	50.4	56.5	48.6	35.7	41.7	42.6	42.9	41.2	37.5	63.5	76.5	75.6	70.6
Tower	89.8	55.6	51.4	52.8	49.1	48.7	44.4	36.7	95.2	55.6	64.8	68.0	57.8
Overall	94.9	80.6	75.0	69.6	70.3	70.6	68.3	62.74	97.0	83.8	91.7	92.2	87.6
Gas Temperatures, °F.													
FDS in	335	330	342	342	355	355	358	355	345	336	338	335	338
FDS out	115	118	118	120	118	122	122	120	112	115	118	118	115
Tower out	100	108	102	110	108	115	118	115	105	105	108	110	106
Liquid temperatures, °F.													
FDS in	98	108	104	110	104	115	118	108	108	110	114	113	110
FDS out	115	118	118	120	118	122	122	118	118	118	118	118	118
Tower in	100	114	118	118	104	120	124	105	118	120	118	123	110
Tower out	115	116	116	120	112	118	122	118	114	115	118	118	114
pH measurements													
Tower outlet	4.2	4.8	4.7	4.8	4.0	4.1	4.1	5.0	4.5	3.6	3.9	4.5	3.8
FDS outlet	11.0	10.9	11.3	11.1	11.2	11.2	11.1	11.2	11.8.	11.7	11.7	11.8	11.8
Hold tank	11.8	11.5	11.3	11.5	11.1	11.3	11.0	11.6	10.8	9.7	11.1	11.4	11.4
Clarifier tank	11.8	11.2	11.1	11.1	11.1	11.0	11.2	11.3	11.4	9.9	11.3	11.2	11.2
Disc Position	1.5	1.0	1.25	0.75	1.0	0.5	1.0	1.75	2.0	2.8	2.4	2.0	2.0
Pressure at outlet, in. Hg gauge					-1.4	-1.6	-1.6	-2.2	-2.6	-3.0	-2.6	-2.4	-2.4

* Efficiency measurements made during these tests were considered in error because of air inleakage at the FDS discharge.

TABLE A-9 cont'd

OPERATING CONDITIONS FOR TASK VIC

Date	4/13	4/13	4/14	4/14	4/14	4/14	4/14	4/14	4/14	4/15	4/15	4/15	4/15	4/15	4/15
Time	1045	2235	1030	1130	1240	1545	1800	1900	2220	1015	1145	1245	1500	1630	1730
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	20	20	20	10	20	20	20	20	20	20
Tower L/G, ratio, gal/mcf	15	15	15	15	15	15	15	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec	150	—	172	173	192	190	192	194	149	148	151	150	103	110	111
Tower pressure drop, inches H_2O	0.9	1.7	1.4	1.6	1.9	1.8	1.9	1.7	1.9	1.4	1.4	1.4	1.4	1.4	1.5
FDS pressure drop, inches H_2O	17.5	18	18.4	18.2	12.5	18.7	19.3	19.8	12	12.5	12.7	13.1	8.4	9.4	10.1
CaO/SO_2 ratio	1.02	1.08	1.04	1.10	1.16	1.00	1.05	1.03	1.07	0.99	0.93	0.78	0.92	0.94	1.03
SO_2 concentration, ppm															
FDS in	1740	1740	1740	1665	1620	1380	1380	1390	1380	1530	1575	2070	2055	2025	1890
FDS out	--	540	780	705	720	600	620	630	735	780	750	945	870	825	780
Tower out	350	0	76	42	32	72**	300**	300**	510**	540	528	552	414	372	334
Fraction of SO_2 removed, %															
FDS	--	69.0	55.2	57.7	55.5	56.5	55.1	54.3	46.7	49.0	52.4	54.3	57.7	59.3	58.7
Tower	--	100	90.3	94.0	95.6	88.0	51.6	52.4	30.61	30.8	29.6	41.6	52.4	54.9	57.2
Overall	79.9	100	95.6	97.5	98.0	94.8	78.3	78.3	63.0	64.7	66.5	73.3	79.9	81.6	82.3
Gas Temperatures, °F.															
FDS in	335	320	315	310	323	315	325	332	330	298	315	308	310	310	318
FDS out	112	112	115	117	118	112	120	118	118	115	115	115	115	115	118
Tower out	103	106	107	109	112	105	110	112	112	113	111	111	110	110	110
Liquid temperatures, °F.															
FDS in	103	-	111	113	115	107	112	-	112	115	115	111	110	110	110
FDS out	117	114	121	120	121	115	118	118	118	119	118	117	115	118	116
Tower in	115	112	116	118	118	109	112	112	112	115	110	112	110	110	112
Tower out	115	112	115	115	117	112	116	114	116	116	115	114	115	115	114
pH measurements															
Tower outlet	2.6	-	4.0	4.5	4.7	4.2	4.6	4.5	4.0	4.2	4.2	4.2	4.4	4.5	4.4
FDS outlet	11.8	11.0	12	12	12.0	11.0	10.8	10.8	10.2	11.1	11.1	10.9	11.2	10.5	10.5
Hold tank	10.8	10.2	11.2	11.2	11.1	11.1	11.2	11.2	11.2	11.0	11.1	11.0	11.1	11.0	10.8
Clarifier tank	11.1	11.0	11.0	11.0	11.0	10.7	10.5	10.5	10.4	10.8	11.1	11.1	11.3	11.3	11.3
Disc Position	1.25*	4.0	1.75	1.75	2.0	2.0	2.0	2.0	1.25	1.37	1.37	1.37	0	.25	.25
Pressure at outlet, in. Hg gauge	-2.7	-2.6	-2.8	-2.8	-2.45	-2.7	-2.7	-2.7	-2.2	-2.1	-2.1	-2.1	-1.9	-2.0	-8.1

* Some scale was deposited on the FDS at this time, subsequent reading is a cleaned disc.

** Water flow to the packed tower to remove the scale build-up.

TABLE A-9 cont'd

OPERATING CONDITIONS FOR TASK VIC

Date	4/15	4/15
Time	2030	2200
Gas Flow, cfm	700	700
FDS L/G ratio, gal/mcf	20	20
Tower L/G, ratio, gal/mcf	15	15
Gas Velocity FDS, ft/sec	112	112
Tower pressure drop, inches H_2O	1.8	1.8
FDS pressure drop, inches H_2O	7.6	8.0
CaO/SO_2 ratio	1.02	1.02
SO_2 concentration, ppm		
FDS in	1800	1800
FDS out	810	810
Tower out	386	396
Fraction of SO_2 removed, %		
FDS	55.0	55.0
Tower	52.3	51.1
Overall	78.6	78.0
Gas Temperatures, °F.		
FDS in	325	325
FDS out	118	118
Tower out	112	112
Liquid temperatures, °F.		
FDS in	110	110
FDS out	118	118
Tower in	112	112
Tower out	114	114
pH measurements		
Tower outlet	4.8	4.8
FDS outlet	11.6	11.7
Hold tank	11.0	11.0
Clarifier tank	11.0	11.0
Disc Position	.25	.25
Pressure at outlet, in. Hg gauge	-1.9	-1.9

TABLE A-10

OPERATING CONDITIONS FOR TASK VI-d

Date	2/25	2/25	2/25	2/25	2/25	2/25	2/25	2/25
Time	0900	1015	1120	1140	1245	1405	1535	1645
Gas Flow, cfm	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	18	18	18	18	18	18	18	18
Tower L/G ratio, gal/mcf	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec	77	77	77	--	77	77	77	77
Tower pressure drop, inches H_2O	0.5	0.5	0.6	--	0.5	0.6	0.6	0.6
FDS pressure drop, inches H_2O	6.0	5.8	5.8	--	5.8	5.6	5.6	6.0
CaO/SO_2 ratio	1.00	1.06	0.98	0.99	1.18	1.0	1.08	0.93
SO_2 concentration, ppm								
FDS in	1920	1935	1950	--	1650	1665	1530	1650
FDS out	--	1110	1110	--	900	900	840	960
Tower out	--	520	660	--	380	380	300	440
Fraction of SO_2 removed, %								
FDS	--	42.5	43.0	--	45.5	45.9	45.1	41.7
Tower	--	53.1	40.6	--	57.7	57.7	64.	54.2
Overall	--	73.1	66.2	--	77.0	77.2	80.4	73.3
Gas temperatures, °F								
FDS in	320	320	320	--	320	315	320	320
FDS out	138	138	144	--	142	138	138	132
Tower out	125	125	134	--	142	137	135	128
Liquid temperatures, °F								
FDS in	122	122	134	--	142	138	135	122
FDS out	138	138	140	--	146	140	138	130
Tower in	138	138	148	--	150	145	110	125
Tower out	135	135	140	--	142	138	135	130
Clarifier	118	118	132	--	140	138	130	115
pH Measurements								
Tower outlet	--	5.8	6.2	--	5.9	6.0	5.9	5.8
FDS outlet	--	10.0	9.0	--	9.8	9.8	11.4	11.3
Hold tank	--	11.0	11.0	--	11.2	11.4	11.4	11.2
Clarifier tank	--	10.0	10.8	--	11.3	11.3	11.4	11.3
Disc position	0	0	0	0	0	0	0	0
Hold tank volume	600	600	600	400	400	400	250	280

* pH on the FDS outlet fluctuated with the variations in flow from the demister. Chart reading was an estimate of pH during high flow conditions.

TABLE A-11

OPERATING CONDITIONS FOR TASK VJ●

Date	5/5 *	5/5 *	5/5 *	5/5 *	5/5	5/5	5/6	5/6	5/6	5/6	5/6	5/6
Time	0930	1030	1115	1330	1500	1600	1045	1220	1320	1500	1600	1700
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	15	15	15	15	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec.	175	176	112	112	113	113	104	127	127	104	104	104
Tower pressure drop, inches H_2O	0.4	0.4	0.4	0.6	0.65	0.7	0.6	0.6	0.6	0.6	0.6	0.6
FDS pressure drop, inches H_2O	6.8	7.3	7.4	7.0	6.4	7.4	6.5	7.8	7.8	7.3	7.5	7.8
CaO/SO_2 ratio	1.12	1.01	1.01	068/108	1.17	1.42	1.03	1.11	1.23	0.92	0.83	1.07
SO_2 Concentration, ppm												
FDS in	1755	1620	1620	2200	2160	1800	2020	1980	1780	1780	1710	1680
FDS out	1065	1050	1020	1320	1110	960	1010	1015	870	930	900	870
Tower out	416	221	266	480	292	130	176	244	209	130	150	130
Fraction of SO_2 removed, %												
FDS	39.3	35.2	37.0	40.0	48.6	46.7	50.0	48.7	51.1	47.8	47.4	48.2
Tower	60.9	79.0	73.9	63.6	73.7	86.5	82.6	76.0	76.0	86.0	83.3	85.1
Overall	76.3	86.4	83.6	78.2	86.5	92.8	91.3	87.7	88.3	92.7	91.2	92.3
Gas Temperature, °F.												
FDS in	328	330	328	330	335	335	322	322	322	320	320	322
FDS out	112	116	116	118	120	120	118	118	118	120	118	118
Tower out	106	108	108	112	115	116	110	110	110	112	112	114
Liquid Temperature, °F.												
FDS in	100	106	105	108	110	110	104	104	102	108	108	108
FDS out	102	110	105	106	105	106	98	105	96	98	94	96
Tower in	96	106	102	104	110	112	102	102	104	106	106	108
Tower out	110	118	115	118	120	120	118	118	118	120	118	118
pH measurements												
Tower outlet	3.6	3.8	3.8	2.4	2.3	-	2.0	2.0	-	2.4	-	2.2
FDS outlet	11.6	11.4	11.4	11.8	11.2	-	11.8	11.4	11.5	11.5	11.3	11.4
Hold tank	11.8	11.5	11.5	11.4	11.5	-	10.3	11.6	11.3	11.3	11.3	11.2
Clarifier tank	11.0	11.0	11.0	11.1	11.2	-	11.4	11.0	11.2	11.3	11.3	11.2
Disc Position	1.75	1.75	0.25	0.25	0.25	0.25	0	0.75	0.75	0	0	0
Pressure at outlet, in. Hg gauge	-1.6	-1.6	-1.6	-2.1	-2.0	-2.0	-1.6	-1.6	-1.6	-1.6	-1.6	-1.6
Slurry Concentration to FDS, weight %	6.7	6.4	6.6	3.4	3.8	3.6	1.1	1.4	1.8	5.8	6.0	6.7

* These tests were not considered at "steady state". 6% slurry tests were repeated at the end of this test series, 5/6/71.

TABLE A-12

OPERATING CONDITIONS FOR TASK VIF

Date	4/16	4/16	4/16	4/16	4/16	4/16	4/16	4/16
Time	1130	1230	1330	1515	1615	1900	2000	2200
Gas Flow, cfm	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10
Tower L/G, ratio, gal/mcf	15	15	15	15	15	15	15	15
Gas Velocity FDS, ft/sec	119	118	118	104	104	104	104	104
Tower pressure drop, inches H_2O	1.6	2.2	2.1	2.9	3.0	3.2	3.0	2.9
FDS pressure drop, inches H_2O	6.7	6.8	7.2	8.2	8.6	6.0	6.0	6.0
CaO/SO_2 ratio	1.23	1.13	1.02	1.05	0.99	1.00	1.60	0.87
SO_2 concentration, ppm								
FDS in	1650	1770	1950	1860	1980	1950	1230	1380
FDS out	600	750	900	840	870	900	570	630
Tower out	324	273	312	48	72	84	120	132
Fraction of SO_2 removed, %								
FDS	63.6	57.6	53.8	54.8	56.0	53.8	53.7	54.3
Tower	46.0	63.6	65.3	94.3	91.7	90.7	78.9	79.0
Overall	80.4	84.6	84.0	97.4	96.4	95.7	90.2	90.4
Gas temperatures, °F.								
FDS in	325	315	315	318	318	322	320	320
FDS out	115	115	115	114	116	118	118	118
Tower out	113	110	115	110	110	118	118	118
Liquid temperatures, °F.								
FDS in	110	110	110	112	112	112	112	112
FDS out	120	120	122	125	124	130	130	130
Tower in	110	110	110	112	112	114	114	114
Tower out	115	115	115	115	116	118	118	118
pH measurements								
Tower outlet	4.4	5.7	5.7	7.0	6.7	5.4	6.5	6.2
FDS outlet	10.7	11.4	11.2	11.5	11.4	11.2	11.4	11.4
Hold tank	11.2	11.4	11.3	11.5	11.4	11.8	11.7	11.6
Clarifier tank	11.2	11.2	11.2	11.2	11.2	10.8	10.9	11.0
Disc Position	0.5	0.5	0.5	0	0	0	0	0
Pressure at outlet, in. Hg gauge	-1.8	-1.8	-1.8	-1.9	-1.9	-2.0	-2.0	-2.0
NaCl Ionic Strength, I	1	1	1	2	2	4	4	4

TABLE A-13

POWER REQUIREMENTS FOR TASKS VIa, b, c, AND d

Task No.	Gas Rate cfm	Δp FDS H_2O"	Δp Tower H_2O"	Liquid Rates			Horsepower (Slurry)			Horsepower (gas)		Total Horsepower	Horsepower/ Megawatt
				FDS gpm	Tower gpm	Clarifier gpm	FDS Pump	Tower Pump	Clarifier Pump	FDS (100% Efficiency)	Tower (100% Efficiency)		
VIa	700	7	.8	12.6	10.5	12.6	0.086	0.092	0.085	0.769	0.087	1.119	3.6
VIb	700	6	.6	12.6	10.5	23.1	0.085	0.092	0.165	0.659	0.0659	1.067	3.4
VIb	700	6	.6	7	10.5	17.5	0.042	0.092	0.124	0.659	0.0659	0.983	3.2
VIb	700	6	.6	16.1	10.5	26.6	0.118	0.092	0.192	0.659	0.0659	1.197	3.8
VIc	700	12	.8	7.0	10.5	17.5	0.042	0.092	0.124	1.31	0.0800	1.648	5.3
VId	700	6	.6	12.6	10.5	23.1	0.085	0.092	0.124	0.659	0.0659	1.026	3.3

TABLE A-14

OPERATING CONDITIONS FOR TASK VIIa

Date	5/10	5/11	5/11	5/11
Time	1745	1015	1140	1500
Gas Flow, cfm	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10
Tower L/G, ratio, gal/mcf	15	15	15	15
Gas Velocity FDS, ft/sec	105	105	105	112
Tower pressure drop, inches H_2O	0.6	0.6	0.6	0.6
FDS pressure drop, inches H_2O	6.8	6.5	8.5	7.1
CaO/SO_2 ratio	0.81	1.0	1.1	1.0
SO_2 concentration, ppm				
FDS in	1845	1860	1610	1500
FDS out	1290	900	750	780
Tower out	471	468	480	325
Fraction of SO_2 removed, %				
FDS	30.1	51.6	53.4	48.0
Tower	63.5	48.0	36.0	58.3
Overall	74.5	74.8	70.2	78.3
Gas Temperatures, °F.				
FDS in	330	330	325	330
FDS out	118	118	117	118
Tower out	110	108	110	112
Liquid temperatures, °F.				
FDS in	108	100	102	100
FDS out	95	100	100	108
Tower in	105	100	102	102
Tower out	118	115	117	108
pH measurements				
Tower outlet	3.4	4.6	4.6	3.9
FDS outlet	5.8	11.6	11.8	11.2
Hold tank	11.0	11.6	11.6	11.5
Clarifier tank	11.1	10.7	6.7	11.0
Disc Position	0	0	0	0.25
Pressure at outlet, in. Hg gauge	-1.4	-1.5	-1.6	-1.6
Slurry Concentration to FDS, % by weight	-	-	22.4	-

TABLE A-15

OPERATING CONDITIONS FOR TASK VIIb

Date	5/13	5/13	5/13	5/13
Time	1400	1530	1630	1730
Gas Flow, cfm	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10
Tower L/G ratio, gal/mcf	20	20	20	20
Gas Velocity FDS, ft/sec.	104	105	105	105
Tower pressure drop, inches H_2O	0.8	0.9	1.0	1.1
FDS pressure drop, inches H_2O	7.1	5.8	6.1	6.1
CaO/SO_2 ratio	0.99	1.0	0.98	0.97
SO_2 concentration, ppm				
FDS in	1860	1830	2040	1890
FDS out	1420	1125	1260	1110
Tower out	297	174	288	156
Fraction of SO_2 removed, %				
FDS	23.7	38.5	38.2	41.3
Tower	79.1	84.5	77.1	85.9
Overall	84.0	90.5	85.9	91.7
Gas Temperature, °F.				
FDS in	322	323	325	323
FDS out	112	113	115	115
Tower out	100	103	106	108
Liquid Temperature, °F.				
FDS in	98	101	103	103
FDS out	92	97	102	102
Tower in	98	102	103	104
Tower out	110	112	115	115
pH measurements				
Tower outlet	3.8	2.5	-	2.2
FDS outlet	5.7	5.8	5.8	5.9
Hold tank	10.6	10.6	10.6	10.4
Clarifier tank	11.1	11.1	11.2	11.2
Disc Position	0	0	0	0
Pressure at outlet, in. Hg gauge	-1.6	-1.6	-1.6	-1.6
Slurry Concentration to FDS, wt. %	15.0	16.0	19.0	19.5
Slurry Concentration to Tower, wt. %	4.4	3.4	3.5	3.4

TABLE A-16

OPERATING CONDITIONS FOR TASK VIIC

Date	5/12	5/12	5/12
Time	1300	1420	1540
Gas Flow, cfm	700	700	700
FDS L/G ratio, gal/mcf	10	10	10
Tower L/G, ratio, gal/mcf	15	15	15
Gas Velocity FDS, ft/sec	105	105	105
Tower pressure drop, inches H_2O	0.7	0.7	0.7
FDS pressure drop, inches H_2O	7.6	6.8	7.5
CaO/SO_2 ratio	0.99	0.985	0.96
SO_2 concentration, ppm			
FDS in	1560	1610	1650
FDS out	1320	1380	1440
Tower out	845	910	1105
Fraction of SO_2 removed, %			
FDS	15.4	14.3	12.7
Tower	36.0	34.1	23.3
Overall	45.8	43.5	33.0
Gas Temperatures, °F.			
FDS in	330	330	328
FDS out	120	120	120
Tower out	113	114	116
Liquid Temperatures, °F.			
FDS in	110	110	110
FDS out	100	102	100
Tower in	109	110	110
Tower out	120	120	120
pH measurements			
Tower outlet	2.1	2.1	2.1
FDS outlet	5.2	5.2	5.2
Hold tank	10.8	10.8	10.6
Clarifier tank	11.1	10.8	10.6
Disc Position	0	0	0
Pressure at outlet, in. Hg guage	-1.6	-1.6	-1.6

TABLE A-17

OPERATING CONDITIONS FOR TASK VIIIC

Date	6/9	6/9	6/9
Time	1730	1830	1915
Gas Flow, cfm	700	700	700
FDS L/G ratio, gal/mcf	10	10	10
Tower L/G ratio, gal/mcf	20	20	20
Gas Velocity FDS, ft/sec.	193	162	162
Tower pressure drop, inches H_2O	0.7	0.7	0.7
FDS pressure drop, inches H_2O	8.2	7.0	7.2
CaO/SO_2 ratio	1.12	1.05	1.08
SO_2 Concentration, ppm			
FDS in	1620	1680	1500
FDS out	510	540	660
Tower out	220	120	100
Fraction of SO_2 removed, %			
FDS	69.0	67.8	56.0
Tower	56.7	77.4	83.0
Overall	86.4	93.0	93.8
Gas Temperature, °F.			
FDS in	310	305	305
FDS out	112	110	110
Tower out	106	108	105
Liquid Temperature, °F.			
FDS in	88	88	88
FDS out	100	102	100
Tower in	102	102	102
Tower out	106	104	106
pH Measurements			
Tower outlet	7.2	7.0	6.6
FDS outlet	9.4	7.4	6.8
Hold tank	8.8	8.5	7.8
Clarifier tank	-	-	-
Disc Position	2.0	1.5	1.5
Pressure at outlet, inches Hg. gauge	-1.6	-1.6	-1.6

TABLE A-18

OPERATING CONDITIONS FOR TASK VIIId

Date	5/18	5/18	5/18
Time	1000	1100	1500
Gas Flow, cfm	700	700	700
FDS L/G ratio, gal/mcf	10	10	10
Tower L/G ratio, gal/mcf	20	20	20
Gas Velocity FDS, ft/sec	105	105	105
Tower pressure drop, inches H_2O	0.8	0.8	0.7
FDS pressure drop, inches H_2O	5.7	6.8	10.8
CaO/SO_2 ratio	1.02	-	0.93
SO_2 Concentration, ppm			
FDS in	1470	1290	1410
FDS out	1200	1140	1200
Tower out	600	570	750
Fraction of SO_2 removed, %			
FDS	18.4	11.6	14.9
Tower	50.0	50.0	37.5
Overall	59.2	55.8	46.8
Gas Temperature, °F.			
FDS in	328	328	330
FDS out	103	104	108
Tower out	103	104	104
Liquid Temperature, °F.			
FDS in	65	65	62
FDS out	100	100	120
Tower in	95	96	98
Tower out	103	104	104
pH measurements			
Tower outlet	2.5	2.5	2.4
FDS outlet	10.4	10.4	6.0
Hold tank	8.6	8.6	6.8
Clarifier tank	10.4	10.4	7.0
Disc Position	0	0	0
Pressure at outlet, in. Hg gauge	-1.5	-1.5	-1.4

TABLE A-19

OPERATING CONDITIONS FOR TASK VIIIe

Date	5/19	5/19	5/19
Time	1700	1815	1900
Gas Flow, cfm	700	700	700
FDS L/G ratio, gal/mcf	10	10	10
Tower L/G ratio, gal/mcf	20	20	20
Gas Velocity FDS, ft/sec.	139	139	139
Tower pressure drop, inches H_2O	1.0	1.0	1.0
FDS pressure drop, inches H_2O	7.0	7.0	7.0
CaO/SO_2 ratio	1.0	0.97	0.96
SO_2 concentration, ppm			
FDS in	1690	1920	2040
FDS out	1110	1330	1620
Tower out	60	96	42
Fraction of SO_2 removed, %			
FDS	34.3	30.7	20.6
Tower	94.6	92.8	97.4
Overall	96.4	95.0	97.9
Gas Temperature, °F.			
FDS in	335	335	335
FDS out	120	120	120
Tower out	118	118	118
Liquid Temperature, °F.			
FDS in	118	118	118
FDS out	110	110	110
Tower in	114	114	116
Tower out	118	118	118
pH measurements			
Tower outlet	5.7	5.6	5.2
FDS outlet	5.5	5.4	5.2
Hold tank	11.0	11.0	10.8
Clarifier tank	-	-	-
Disc Position	1.0	1.0	1.0
Pressure at outlet, in. Hg gauge	-1.6	-1.6	-1.6
Slurry Concentration to FDS weight %	4	4	4

APPENDIX B

OPERATING CONDITIONS AND RESULTS
For
THE LIMESTONE TESTS

TABLE B-1

Limestone Materials Received From TVA

Type	Particle Size	Amount Received, lb	Chemical Analysis, %				
			CaO	MgO	K_2O	Na_2O	Ign Loss
Tiftona Limestone	75%-200M	4700	50.5	1.5	0.4	0.3	41.5
Cement Kiln Dust	90%-200M	1000	41.5	2.4	3.1	0.19	22.8
Selma Chalk	89%-200M	1000	43.1	0.59	0.49	0.18	35.1
Hydrated Lime	99%-325M	1000	70.2	--	--	--	--
Tiftona Limestone	89%-325M	350	50.5	1.5	0.4	0.3	41.5

TABLE B-2

Limestone Test Program

A 1. Alter existing pilot plant hardware arrangement to accommodate the mode shown in Figure IV-21. This involves relocating pumps, lime feeder, piping, etc.

A 2. Measure efficiency of SO_2 removal when scrubbing with a 2% chalk slurry. Mode of operation as per Figure IV-21.
L/G to Venturi to be 10 gal/1,000 cf, and to packed tower to be 40 gal/1,000 cf. Pressure drop in venturi to be 7".

A 3. Measure the efficiency of SO_2 removal when scrubbing with a 2% cement dust slurry. Mode of operation and operating parameters as per Task A2.

A 4. Measure the efficiency of SO_2 removal when scrubbing with a 2% slurry of "local" limestone having a mesh size of 90% minus 325. Mode of operation and operating parameters as per Task A2.

A 5. Measure the efficiency of SO_2 removal when scrubbing with a 2% slurry of "local" limestone having a particle size of 70% minus 200 mesh. Mode of operation and operating parameters as per Task A2.

A 6. Measure the SO_2 removal efficiency when employing the slurry and operating parameters as in Task A5. The mode will be altered such that the slurry goes only to the packed tower and fresh water will be delivered to the venturi.

A 7. Measure the SO_2 removal efficiency when employing the mode and slurry as in Task A6. The liquor rate to the packed tower

TABLE B-2 cont'd

will be reduced to 20 gal. per 1,000 cf. All other operating parameters as per Task A6.

A 8. Measure the SO_2 removal efficiency when employing the mode and operating parameters as in Task A5 but increasing the stoichiometric feed rate to 150% of the stoichiometric requirement.

A 9. Measure the SO_2 removal efficiency when scrubbing with a 2% slurry of lime (CaO). The mode and operating parameters as per Task A2.

A 10. Return pilot plant hardware to the arrangement existing prior to this test program. Open critical elements for inspection and clean as necessary.

A 11. Final Report Preparation and Presentation:
Minimal data analysis will be employed. Tabulation and graphical data representations will be incorporated into a single comprehensive report.

TABLE B-3

PROPOSED TEST PLAN

Task	
C 1	Alter pilot plant equipment so that the recirculation system shown in Figure IV-23 is established. Items of work needed for the hardware change are: (a) change piping to include clarifier in the circuit, (b) install limestone feeder on hold tank, (c) fabricate and install helical coil for the hold tank, and (d) set up sampling equipment for particulate analysis.
C 2	A continuous "around the clock" plugging study for a minimum of 40 hours will be performed. L/G to the venturi will be 10 gal/1,000 cf and the packed tower will be 30 gal/1,000 cf. Gas flow rate to be such that a velocity of 9.5 ft. per sec. will be maintained. Pressure drop on the FDS to be controlled at 7 inches of water. A 5% w/w slurry is planned for the packed tower scrubbing liquor. Hold volume for the packed tower recirculation tank will allow 30 minutes residence time for slurry. Following this continuous test, remove the packing from the tower and compare the weight of each section of packing to its original. Flooded disc section will also be inspected for scaling.
C 3	Repeat Task C2 with the L/G for the packed tower at 50 gal/1,000 cf. Packing will be removed and weighed after 40 hours of continuous operation.

TABLE B-3 cont'd

Task	
C 4	Repeat Task C3 with a slurry residence time of 10 minutes in the hold tank.
C 5	Repeat Task C2 with a slurry residence time of 10 minutes in the hold tank.
C 6	From Task C2 through C5, the minimum scaling mode will be determined and a continuous test will be performed for at least 96 hours. Packing will be removed and weighed after the 96 hours of operation.
C 7	Pilot hardware will be returned to the arrangement needed for the original contract work. Recondition pumps if packing and impellers are damaged during the high slurry test program.
C 8	A summary report will be prepared which will state the results of this work with minimal data analysis.

TABLE B-4

OPERATING CONDITIONS FOR TASK C-2

Date	1/13	1/14	1/14	1/14	1/14	1/14	1/14	1/14	1/14	1/15	1/15
Time	2130	0100	0300	0600	1000	1200	1420	2030	2300	0100	0330
Limestone Used*	B	B	B	B	B	B	B	B	B	B	B
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	30	30	30	30	30	30	30	30	30	30	30
Gas velocity FDS, ft/sec. ***	110	115	130	137	108	--	110	115	116	98	88
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	1.2	2.2	2.2	3.6	4.3	6.0	7.0	8	6.8	6.0	13.2
FDS pressure drop, inches H_2O	6.9	7.0	7.0	7.0	7.0	6.8	8.0	7.0	6.9	7.0	7.0
CaO/SO_2 ratio (inlet analysis)	1	0.6	1	1	0.9	--	1	1.7**	1.0	1.1	1.1
SO_2 Concentration, ppm											
FDS in ****	1350	1580	1500	1550	1930	1500	1325	1032	1063	1030	1030
FDS out	1525	1600	1550	1230	--	1560	1500	1225	1250	875	950
Tower out	695	1020	880	850	1050	660	336	520	570	320	460
Fraction of SO_2 removed, %											
FDS	--	--	--	--	--	--	--	--	--	15.1	8.3
Tower	54.5	36.3	43.3	30.9	--	57.7	77.6	57.6	54.4	63.5	51.6
Overall	--	--	--	--	45.6	--	--	--	--	69.9	55.4
Gas temperatures, °F											
FDS in	329	342	350	350	349	--	345	340	335	338	335
FDS out	118	122	122	120	122	--	122	120	122	120	110
Tower out	111	110	112	112	--	--	115	116	118	110	102
Liquid temperatures, °F											
FDS in	110	122	122	122	110	--	105	116	115	112	110
FDS out	115	122	120	120	122	--	121	120	120	120	115
Tower in	115	122	120	115	119	--	118	118	120	115	108
Tower out	115	120	118	120	122	--	120	118	118	118	110
PH measurements											
Tower outlet	4.8	4.6	4.8	4.6	4.8	--	5.2	5.6	5.2	4.9	4.2
FDS outlet	6.3	6.2	6.4	6.5	6.4	--	5.3	6.6	6.6	6.5	6.4
Hold tank	5.3	5.0	5.2	5.2	5.9	--	5.5	5.2	5.5	5.6	5.4
Clarifier tank	4.7	4.6	4.6	4.6	4.8	--	4.7	4.7	4.7	4.7	4.6

*Limestone received from TVA for the "around-the-clock" tests designated as A; Old Tiftona from TVA shown as Type B.

**Stoichiometry was 1.0 at 2015

***FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-3

Date	1/11	1/12	1/12	1/12	1/13
Time	1600	0930	1730	2230	0100
Limestone Used*	A	A	A	A	A
Gas Flow, cfm	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10.0	10.3	11	--
Tower L/G ratio, gal/mcf	50	45.5	61.0	40	--
Gas velocity FDS, ft/sec.***	76	76	76	76	76
Gas velocity tower, ft/sec.	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	1.0	1.0	1.0	1.2	1.7
FDS pressure drop, inches H_2O	--	5	4.9	4.4	4.4
CaO/SO_2 ratio (inlet analysis)	1	1	1	1	1
SO_2 concentration, ppm					
FDS in****	--	1350	1825	1920	2000
FDS out	--	500	1785	733	--
Tower out	--	150	625	429	500
Fraction of SO_2 removed, %					
FDS	--	63.0	2.2	61.9	--
Tower	--	60.0	65.0	41.5	--
Overall	--	88.9	65.8	77.7	75.0
Gas temperatures, °F					
FDS in	--	330	342	364	342
FDS out	--	125	122	122	122
Tower out	--	127	110	121	118
Liquid temperatures, °F					
FDS in	--	130	122	122	122
FDS out	--	122	122	122	122
Tower in	--	12	118	121	120
Tower out	--	125	121	122	122
pH measurements					
Tower outlet	--	6.0	5.2	5.6	5.3
FDS outlet	--	4.0	4.4	4.1	3.8
Hold tank	--	6.3	5.6	5.7	5.4
Clarifier tank	--	--	5.2	5.1	5.0
Hold tank volume	600	600	760	700	700

*Limestone received from TVA for the "around-the-clock" tests designated as A Old Tiftona from TVA shown as Type B.

***FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test. High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-4

Date	1/18	1/18	1/18	1/18	1/18	1/19	1/19	1/19	1/19	1/19	1/19
Time	1330	1700	1900	2130	2330	0130	0530	1030	1440	1600	2230
Limestone Used	B	B	B	B	B	B	B	B	B	B	B
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	40	40	40	40	40	40	40	40	40	40	40
Gas velocity FDS, ft/sec. ***	138	138	138	138	138	138	138	138	138	138	138
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	--	0.8	0.5	0.7	0.7	0.7	0.7	0.8	1.4	1.2	0.9
FDS pressure drop, inches H_2O	--	7.8	6.4	7.0	7.0	7.0	7.0	7.8	7.9	7.2	7.0
CaO/SO_2 ratio	--	1.6	1.0	.9	1.0	0.9	1.2	0.9	0.9	1.4	--
SO_2 Concentration, ppm											
FDS in ****	--	1150	1750	1700	1325	1230	950	1250	--	1350	1475
FDS out	--	1390	1900	1900	1500	1250	1025	1500	--	1350	1400
Tower out	--	400	150	1125	320	350	488	--	700	680	800
Fraction of SO_2 removed, %											
FDS	--	--	--	--	--	--	--	--	--	--	5.1
Tower	--	71.3	92.2	40.8	78.7	72.0	52.4	--	--	49.7	42.9
Overall	--	--	--	--	--	--	--	--	--	--	45.8
Gas temperatures, °F											
FDS in	--	335	335	330	330	330	330	335	330	333	338
FDS out	--	110	90	105	105	105	105	115	110	112	112
Tower out	--	105	80	100	100	100	102	105	105	105	105
Liquid temperatures, °F											
FDS in	--	112	100	105	105	105	105	112	110	112	108
FDS out	--	125	120	115	110	110	112	120	118	120	120
Tower in	--	108	83	100	100	100	100	105	105	105	105
Tower out	--	111	100	110	-	110	110	-	-	112	-
pH measurements											
Tower outlet	--	4.9	5.7	5.5	5.2	5.3	4.2	4.7	4.9	4.9	4.8
FDS outlet	--	6.6	6.9	7.0	5.4	7.2	7.4	6.7	6.3	6.7	6.5
Hold tank	--	5.6	6.3	5.7	5.7	5.9	5.8	5.7	5.5	5.5	5.5
Clarifier tank	--	7.1	7.1	7.1	7.1	7.2	-	-	-	-	-

***FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test. High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-4

Date	1/19	1/20	1/20
Time	2230	0330	0330
Limestone Used	B	B	a
Gas Flow, cfm	700	700	700
FDS L/G ratio, gal/mcf	10	10	10
Tower L/G ratio, gal/mcf	40	40	40
Gas velocity FDS, ft/sec.***	138	138	138
Gas velocity Tower, ft/sec.	8.35	8.35	8.35
Tower pressure drop, inches H_2O	1.6	1.6	1.6
FDS pressure drop, inches H_2O	7.0	7.0	7.0
CaO/SO_2 ratio	--	1.0	--
SO_2 Concentration, ppm			
FDS in ****	1650	850	500
FDS out	1625	702	425
Tower out	1040	400	100
Fraction of SO_2 removed, %			
FDS	1.6	17.4	15.0
Tower	36.0	42.4	76.5
Overall	37.0	52.9	80.0
Gas temperatures, °F			
FDS in	325	332	325
FDS out	110	105	100
Tower out	103	104	98
Liquid temperatures, °F			
FDS in	105	102	102
FDS out	120	112	112
Tower in	105	102	102
Tower out	-	102	-
pH measurements			
Tower outlet	4.7	4.6	-
FDS outlet	7.1	6.4	-
Hold tank	5.3	5.4	-
Clarifier tank	-	-	-

***FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test. High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-5

Date	1/20	1/20	1/20	1/20	1/20	1/20	1/20	1/21	1/21	1/21	1/21	1/21
Time	1130	1230	1400	1700	1900	2100	2300	0100	0300	0500	0730	0915
Limestone Used	B	B	B	B	B	B	B	B	B	B	B	B
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	40	40	40	40	40	40	40	40	40	40	40	40
Gas velocity FDS, ft/sec.***	141	141	141	141	141	141	141	141	115	115	138	108
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	0.8	1.0	1.0	0.8	1.1	1.1	1.1	1.1	1.9	1.9	3.1	5.4
FDS pressure drop, inches H_2O	6.6	6.4	7.2	6.6	7.2	6.7	7.0	7.0	7.0	7.0	7.0	5.8
CaO/SO_2 ratio	--	--	1.1	1.1	2.1	1.4	1.1	2.0	1.7	1.45	--	1.45
SO_2 Concentration, ppm												
FDS in ****	1250	1075	1050	950	850	750	850	775	600	500	600	750
FDS out	800	600	425	350	300	250	300	378	250	234	300	275
Tower out	17	200	235	100	185	155	285	330	160	125	150	120
Fraction of SO_2 removed, %												
FDS	36.0	44.2	59.5	63.1	64.7	66.6	64.7	51.3	58.3	53.2	50.0	63.3
Tower	97.8	66.7	44.7	71.4	38.3	38.0	5.00	12.7	36.0	53.4	50.0	54.6
Overall	98.6	81.4	77.6	89.4	78.3	79.3	66.5	57.4	73.4	75.0	75.0	84.0
Gas temperatures, °F												
FDS in	335	330	330	330	333	332	328	330	332	332	--	335
FDS out	85	108	110	105	115	115	117	118	112	110	--	115
Tower out	70	102	105	102	112	112	113	112	112	105	--	110
Liquid temperatures, °F												
FDS in	100	105	105	105	108	111	112	110	108	102	--	85
FDS out	110	115	115	115	118	118	118	118	118	126	--	--
Tower in	75	102	105	100	112	112	115	115	112	106	--	110
Tower out	80	105	110	105	115	115	118	118	118	-	--	--
pH measurements												
Tower outlet	5.5	5.0	5.2	4.8	5.1	4.5	5.3	4.6	5.0	4.8	--	4.8
FDS outlet	5.0	6.0	4.9	5.2	4.9	4.9	4.8	4.8	4.8	4.7	--	4.3
Hold tank	6.3	7.1	6.2	6.4	6.6	6.4	6.4	6.2	6.4	6.0	--	5.7
Clarifier tank	--	--	--	--	6.4	5.9	5.7	5.6	5.5	5.5	--	4.3

***FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test. High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-5

Date	1/21	1/21	1/21	1/21	1/21	1/21	1/21
Time	1100	1300	1500	1700	1900	2100	2300
Limestone Used	B	B	B	B	B	B	B
Gas Flow, cfm	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	40	40	40	40	40	40	40
Gas velocity FDS, ft/sec.***	111	132	115	115	115	115	115
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	5.0	5.0	3.2	3.3	3.6	3.9	4.3
FDS pressure drop, inches H_2O	5.8	6.0	5.2	4.6	4.2	4.6	4.0
CaO/SO_2 ratio	1.27	1.3	0.8	1.0	0.9	0.9	0.9
SO_2 Concentration, ppm							
FDS in ****	825	890	1125	1300	1500	1600	1650
FDS out	225	225	1050	1250	1375	1550	1525
Tower out	210	240	330	370	355	425	360
Fraction of SO_2 removed, %							
FDS	72.8	74.8	7.7	5.7	8.4	3.1	7.6
Tower	6.7	--	68.6	70.4	74.2	72.6	76.4
Overall	74.6	73.1	70.7	71.6	76.4	73.5	78.2
Gas temperatures, °F							
FDS in	335	340	335	335	335	335	333
FDS out	120	115	120	118	119	121	120
Tower out	115	70	120	119	119	120	120
Liquid temperatures, °F							
FDS in	110	105	118	114	115	118	118
FDS out	-	115	120	118	118	120	119
Tower in	120	115	120	119	119	121	121
Tower out	-	--	--	--	--	--	--
pH measurements							
Tower outlet	4.9	4.8	3.0	3.8	5.9	5.8	5.3
FDS outlet	4.5	3.7	3.7	4.1	4.6	4.9	5.1
Hold tank	5.7	5.4	5.7	5.8	6.2	6.0	5.9
Clarifier tank	5.2	5.1	5.0	4.8	4.2	4.2	4.2

*** FDS gas velocity based on outlet gas volume.

**** Some difficulties in SO_2 analyses were experienced during this test. High efficiencies and negative absorption on FDS should not be considered valid.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-6

Date	1/25	1/25	1/25	1/26	1/26	1/26	1/26	1/26	1/26	1/26	1/27
Time	1830	2130	2400	0200	0400	1115	1230	1500	1715	2230	0300
Limestone Used	B	B	B	B	B	B	B	B	B	B	B
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	45	45	45	45	45	45	45	45	45	45	45
Gas velocity FDS, ft/sec.***	127	141	141	141	141	141	135	135	135	141	141
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	1.0	1.0	1.1	0.9	1.0	1.0	1.0	1.0	0.9	0.9	0.9
FDS pressure drop, inches H_2O	6.4	7.0	6.5	6.8	7.3	6.0	6.2	5.8	6.2	7.0	7.0
CaO/SO_2 ratio	0.9	1.0	0.9	1.0	1.0	1.0	1.2	1.2	0.9	1.1	0.9
SO_2 Concentration, ppm											
FDS in	1700	1700	1920	1930	1890	1680	1360	1260	1380	1400	1565
FDS out	1580	1635	1900	1760	1735	1560	1220	1160	--	1320	1440
Tower out	540	620	880	560	460	320	100	34	--	240	280
Fraction of SO_2 removed, %											
FDS	7.1	3.9	1.0	8.8	8.2	7.2	10.3	8.0	--	5.8	8.0
Tower	65.9	62.1	53.7	68.2	73.5	79.5	91.8	97.1	--	81.9	80.6
Overall	68.3	63.5	54.2	71.0	75.7	81.0	92.7	97.4	--	82.9	82.2
Gas temperatures, °F											
FDS in	335	328	335	330	330	--	--	330	335	333	335
FDS out	105	120	120	118	120	118	118	118	118	118	115
Tower out	108	120	120	118	120	118	118	118	115	115	116
Liquid temperatures, °F											
FDS in	40	114	112	115	118	112	112	112	110	114	114
FDS out	85	118	120	118	120	120	120	120	118	118	118
Tower in	108	118	121	118	120	118	118	118	115	115	114
Tower out	108	118	121	120	122	118	120	120	118	118	118
pH measurements											
Tower outlet	5.4	5.4	4.9	5.2	5.4	5.5	5.9	6.6	4.3	5.7	4.8
FDS outlet	4.6	4.4	5.2	5.0	4.8	5.4	6.8	6.8	6.6	4.8	4.8
Hold tank	5.3	5.5	5.3	5.7	5.4	6.2	6.4	6.1	6.0	6.4	6.3
Clarifier tank	3.2	3.4	3.4	4.7	4.6	4.3	4.5	4.9	4.9	4.6	4.5

***FDS gas velocity based on outlet gas volume.

TABLE B-4 cont'd

OPERATING CONDITIONS FOR TASK C-6

Date	1/29	1/29	1/29	1/30	1/30	1/30	1/30	1/30	1/30	1/30	1/30	1/30	1/31
Time	1405	1620	2100	0200	0600	1230	1330	1415	1505	1830	2100	2330	0100
Limestone Used	A	A	A	A	A	A	A	A	A	A	A	A	A
Gas Flow, cfm	700	700	700	700	700	700	700	700	700	700	700	700	700
FDS L/G ratio, gal/mcf	10	10	10	10	10	10	10	10	10	10	10	10	10
Tower L/G ratio, gal/mcf	45	45	45	45	45	45	45	45	45	45	45	45	45
Gas velocity FDS, ft/sec.***	127	127	127	141	141	151	141	141	141	141	141	141	141
Gas velocity Tower, ft/sec.	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35	8.35
Tower pressure drop, inches H_2O	0.8	0.8	0.8	0.7	0.6	0.6	--	--	0.6	1.1	1.1	0.9	0.9
FDS pressure drop, inches H_2O	6.4	7.0	7.0	7.0	7.0	6.4	--	--	6.0	7.2	7.6	6.4	7.0
CaO/SO_2 ratio	1.2	1.2	1.2	1.2	1.1	1.3	1.2	1.2	1.2	1.2	1.2	1.2	1.2
SO_2 Concentration, ppm													
FDS in	1440	1440	1360	1280	1480	1450	1420	1240	1260	1160	1000	1000	1080
FDS out	1320	1340	1280	1140	1380	1340	1300	--	1120	1100	960	979	1000
Tower out	360	220	210	280	280	400	320	120	140	84	20	20	72
Fraction of SO_2 removed, %													
FDS	8.4	8.0	5.9	11.0	6.8	7.6	8.5	--	11.2	5.2	4.0	4.0	2.9
Tower	72.8	83.5	83.6	75.5	79.8	70.2	75.4	--	87.5	92.4	97.9	97.9	92.8
Overall	75.0	84.8	84.6	78.2	81.1	72.5	77.5	90.4	88.9	92.8	98.0	98.0	93.3
Gas temperature, °F													
FDS in	342	340	342	335	330	325	--	--	330	330	320	325	330
FDS out	122	112	122	120	120	120	--	--	120	116	120	118	118
Tower out	122	110	120	118	118	120	--	--	118	112	118	112	112
Liquid temperatures, °F													
FDS in	119	100	118	118	116	118	--	--	115	115	114	115	116
FDS out	126	118	125	122	122	122	--	--	122	122	120	121	120
Tower in	122	112	121	118	118	120	--	--	118	113	120	112	114
Tower out	125	115	123	120	120	122	--	--	122	118	120	118	120
pH measurements													
Tower outlet	5.8	3.5	5.8	5.6	5.4	5.4	--	--	5.5	5.1	5.3	4.4	4.8
FDS outlet	5.4	4.2	5.4	5.4	4.7	5.5	--	--	6.0	4.4	4.4	4.0	4.9
Hold tank	7.6	5.4	6.4	6.2	6.2	5.8	--	--	6.2	6.4	6.3	6.2	6.0
Clarifier tank	3.6	5.0	5.0	5.2	4.9	5.2	--	--	5.2	5.1	4.9	5.0	5.0

***FDS gas velocity based on outlet gas volume.

TABLE B-5

Limestone Materials Supplied By TVA For The Scaling Test Series*

Designation	Shipment A	Shipment B
Particle Size	61%-200 Mesh	75% - 200 Mesh
Amount Received, lbs.	2000	4700
Chemical Analysis		
CaO %	50.8	50.5
MgO %	-	1.5
K_2O %	-	0.4
Na_2O %	-	0.3
Ign. Loss	41.5	41.5

* Limestone supplied by TVA called "Tiftona Limestone". Shipment "A" was a special supply of $CaCO_3$ for the scaling test series; Material "B" was received from TVA for the first test series.

TABLE B-6

Analytical Results on Solids from TIDD Pilot-Plant Limestone Slurry Scrubbing Tests—January 13-31, 1971

Task No.,[a] date, end time	Limestone slurry concentration and source[b]	Sample			Stack gas		SO_2 removed, % of input (cumulative)	Chemical analysis, of solids, %					CaO utilization, %	Solids composition calculated from chemical analysis				
		Point	% solids	Filtrate pH	Flow rate ft^3/min.	Analysis, ppm SO_2		CaO	S Total	S SO_2	S SO_3	Acid insol.		$CaCO_3$	$CaSO_4 \cdot 2H_2O$	$CaSO_3 \cdot 1/2H_2O$	Acid insol.	Total
C-2 1/13/71 1700 hr.	2%, TL-3	Hold tank	3.0	3.1	700	1350[c]	-	42.6	11.0	1.9	9.1	4.2	45.2	42	49	8	4	103
		PT outlet	3.5	2.0			49	41.4	9.6	2.0	7.6	6.1	40.6	44	41	8	6	99
		FDS outlet	57.5	7.6			49	43.7	2.0	0.4	1.6	9.4	8.0	71	9	2	9	91
C-2 1/14/71 2315 hr.	2%, TL-3	Hold tank	1.2	2.3	700	1065	-	37.8	17.3	3.5	13.8	5.6	80.1	14	74	14	6	108
		PT outlet	1.4	2.2			46	35.3	14.2	1.4	12.8	7.9	70.5	20	69	6	8	103
		FDS outlet	3.7	2.0			46	10.7	5.0	0.3	4.7	54.3	82.2	3	25	1	54	83
C-2 1/15/71 0330 hr.	2%, TL-3	Hold tank	5.3	3.2	700	1030	-	20.6	8.2	0.2	8.0	37.5	69.9	11	43	1	38	93
		PT outlet	7.3	2.7			47	15.8	7.7	0.2	7.5	44.1	85.4	4	40	1	44	89
		FDS outlet	8.6	2.1			55	14.3	6.7	0.3	6.4	47.7	81.8	5	34	1	48	88
C-3 1/12/71 1900 hr.	5%, TL-4	Hold tank	4.7	7.8	700	1825	-	42.0	11.1	6.6	4.5	6.4	46.2	40	24	27	6	97
		PT outlet	4.4	4.3			64	37.0	12.9	8.2	4.7	12.6	61.1	26	25	33	13	97
		FDS outlet	5.6	2.2			66	20.0	9.6	6.2	3.4	38.2	84.0	6	18	25	38	87
C-5 1/21/71 0700 hr.	5%, TL-4	Hold tank	1.1	7.2	700	600	-	36.4	9.8	0.12	9.68	5.9	47.2	34	52	1	6	93
		PT outlet	2.4	6.6			25	35.8	3.3	0.24	3.06	11.5	16.2	54	16	1	12	83
		FDS outlet	1.6	6.2			75	11.0	5.3	0.14	5.16	48.5	84.5	3	28	1	49	81
C-6 1/25/71 2230 hr.	5%, TL-3	Hold tank	9.0	7.5	700	1700	-	35.4	13.5	0.80	12.7	3.6	66.7	21	68	3	4	96
		PT outlet	7.9	6.2			60	31.0	14.5	0.83	13.67	7.6	81.9	10	74	3	8	95
		FDS outlet	27.1	4.8			64	18.4	6.8	0.23	6.57	35.7	64.7	12	35	1	36	84
C-6 1/26/71 1300 hr.	5%, TL-3	Hold tank	5.4	7.5	700	1360	-	32.6	15.5	0.18	15.32	4.4	83.1	10	82	1	4	97
		PT outlet	3.9	7.3			82	31.3	15.7	0.30	15.4	5.3	87.9	7	83	1	5	96
		FDS outlet	21.8	4.2			92	24.1	10.5	1.3	9.2	27.3	76.3	10	50	5	27	92
C-6 1/26/71 2100 hr.	5%, TL-3	Hold tank	7.2	7.4	700	1400	-	34.1	15.0	0.2	14.8	3.9	77.1	14	80	1	4	99
		PT outlet	6.4	6.3			77	28.0	14.5	0.7	13.8	12.5	90.7	5	74	3	13	95
		FDS outlet	27.8	5.0			83	15.7	7.8	1.6	6.2	41.7	87.3	4	33	6	42	85
C-6 1/29/71 1400 hr.	5%, TL-4	Hold tank	6.0	7.3	700	1440	-	30.3	15.4	0.18	15.22	7.1	89.1	6	82	1	7	96
		PT outlet	4.8	7.1			67	29.6	15.5	0.31	15.19	8.0	91.6	5	82	1	8	96
		FDS outlet	17.3	5.2			75	22.4	11.2	0.33	10.87	24.5	87.5	5	58	1	25	89
C-6 1/29/71 2100 hr.	5%, TL-4	Hold tank	8.3	6.8	700	1360	-	36.3	12.3	0.19	12.11	4.1	59.2	26	65	1	4	96
		PT outlet	13.7	6.1			79	31.1	15.5	0.47	15.03	5.1	87.1	7	81	2	5	95
		FDS outlet	14.2	5.5			85	17.3	8.3	0.19	8.11	34.5	83.8	5	44	1	35	85
C-6 1/30/71 1230 hr.	5%, TL-4	Hold tank	5.5	6.2	700	1450	-	36.8	11.6	0.73	10.87	4.6	55.2	30	58	3	5	96
		PT outlet	4.5	6.0			65	34.4	13.7	0.75	12.95	4.6	69.8	19	70	3	5	97
		FDS outlet	9.2	4.1			72	28.1	9.0	1.2	7.8	20.4	56.2	4	42	5	20	71
C-6 1/31/71 0105 hr.	5%, TL-4	Hold tank	8.0	7.3	700	1080	-	34.6	13.4	0.3	13.1	3.4	67.9	20	70	1	3	94
		PT outlet	5.9	6.8			86	31.4	15.6	0.3	15.3	4.2	86.9	7	82	1	4	94
		FDS outlet	16.1	5.3			93	19.5	8.9	1.2	7.7	33.3	79.6	7	41	5	33	86

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[a] All samples taken during closed-loop operation; Tiftona limestone used in all tests.
[b] TL-3 = TVA ground material, 75% -200 mesh, 50.5% CaO; TL-4 = commercially ground material, 61% -200 mesh, 50.8% CaO.
[c] Gas analysis taken 4.5 hours after sample taken.